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Fluidized Bed Combustion Effectiveness of an SO2 Control Technology for Industrial Boilers

Radian Corp., Research Triangle Park, NC

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16. ABSTRACT

Atmospheric fluidized bed combustion (AFBC) boilers have developed rapidly over recent years and are now offered commercially in several different configurations. SO_2 reduction levels of 90 percent and above have been achieved by coal-fired AFBC boilers in the industrial size category. Based on the data available, industrial FBC NO_x emissions have been consistently below 0.5 lb/million Btu. PM emissions of less than 0.5 lb/million Btu have been routinely achieved with fabric filters. AFBC boiler system costs were compared with costs for a conventional boiler equipped with an FGD system and with costs for a conventional boiler using low sulfur compliance coal. The conclusions drawn from the economic analyses are that (1) studied cost difference between AFBC Technology, conventional boiler/FGD systems, and compliance coal combustion are projected to be small over the SO_2 emission range of 1.7 to 0.8 lb/million Btu and SO_2 reduction range of 65 to 90 percent, and (2) that cost competitiveness among these technologies is not expected to change significantly as the emission limitations change over this range. Absolute economic competitiveness among these options will be sensitive to site-specific parameters and decided on a case-by-case basis.

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SECTION 1
INTRODUCTION

Fluidized bed combustion (FBC) of coal is now considered a viable alternative for industrial steam generation. Several vendors are offering industrial FBC steam generators on a commercial basis. Competing with FBC technology are two other options for burning coal in the industrial setting and meeting applicable emission limits: conventional boilers equipped with flue gas desulfurization (FGD) systems; and uncontrolled combustion of low-sulfur, or "compliance", coals.

The U. S. Environmental Protection Agency (EPA) is currently involved in the revision of sulfur dioxide (SO_2) New Source Performance Standards (NSPS) for industrial boilers. The overall objective of this study is to evaluate the development status of FBC systems and the influence of alternative SO_2 emission limits on the economic competitiveness of FBC relative to the two competing SO_2 control options. This overall objective has been expanded into three specific sub-objectives:

1. To update the FBC technology status information and emissions data appearing in the FBC Integrated Technology Assessment Report (ITAR) of November 1979.[1] The emphasis of this update will be on SO_2 emissions but nitrogen oxide (NO_x) and particulate matter (PM) emissions will also be considered;

2. To evaluate the economic competitiveness of FBC technology relative to the two competing SO_2 control options and determine how this competitiveness would be affected by alternative emission limits;

3. To determine under what conditions, if any, FBC technology would be economically favored over the two competing control options.

In writing this report, we have assumed that the reader is familiar with the FBC-ITAR. This report deals primarily with the changes that have occurred in the technology and emissions/performance data since the date of the ITAR. Although this report will build on the ITAR, it is intended to serve as a stand-alone document. Therefore, if material is covered adequately in the ITAR, that discussion is only summarized here; the emphasis in this report is on new information not covered in the ITAR.

This work was performed from May 1983 to September 1984 under the direction of the Office of Policy and Resource Management, EPA with consultation from the Office of Air Quality Planning and Standards, EPA.

Section 2 of this report contains an Executive Summary of the study's findings. An evaluation of the development status of FBC technology is presented in Section 3. Emissions and performance data related to both SO_2 control and NO_x and particulate matter (PM) control are discussed in Section 4. Section 5 describes the development of the FBC cost algorithm and compares algorithm projections with independent vendor cost estimates. FBC cost competitiveness relative to conventional boiler/FGD systems and compliance coal use as a function of SO_2 emission limits is evaluated in Section 6.

1.1 REFERENCES

1. Young, C. W., et al. (GCA Corporation). Technology Assessment Report for Industrial Boiler Applications: Fluidized-Bed Combustion. United States Environmental Protection Agency Report No. EPA-600/7-79-178e. November 1979.

SECTION 2
EXECUTIVE SUMMARY

The major objectives of this study are to (1) update the FBC technology status and emissions data since the time of the FBC-ITAR and (2) develop an economic comparison of FBC technology with conventional boiler/FGD systems and compliance coal combustion for industrial boilers operating under a range of SO_2 emission control levels. While the primary emphasis of this investigation is on the SO_2 control capabilities of FBC technology, NO_x emisions, PM emissions, and boiler performance parameters have also been examined.

Commercial Availability

Atmospheric fluidized bed combustion (AFBC) boilers have developed rapidly over the past four years and are now offered commercially in several different configurations. Design alternatives which are currently available include the conventional bubbling fluidized bed (with or without solids recycle), staged fluidized beds, circulatory fluidized beds, and staging of combustion air (for NO_x control). Pressurized FBC technology has been under development for several years, but it is not a likely candidate for commercial applications in the industrial boiler segment except for very large-scale industrial boilers. Pressurized FBC boilers are not considered further in this study.

Of the 36 manufacturers offering AFBC boilers on a commercial basis, 20 are located in the U. S. The domestic manufacturers offer units ranging in size from 2,000 to 600,000 lb/hr of steam at conditions up to 2650 psig and 1050°F (2.3 to 935 million Btu/hr heat input.) Many vendors offer system guarantees covering performance in such areas as steam quality and quantity, emissions, and combustion efficiency. A majority of the existing and planned units in the U. S. and abroad are based on the conventional bubbling bed design; a few units incorporate the circulating bed design; and only two units have staged beds. Fuel feedstocks vary widely for these units from low rank fuels (e.g., lignite, peat, agricultural and municipal wastes) to coal, oil, and natural gas. Many units are designed to burn multiple fuels, either separately or in combination. This fuel feedstock flexibility is an advantage that FBC boilers enjoy over conventional boilers as a result of their high thermal inertia. FBC and conventional boiler/FGD systems demonstrate similar performance with respect to boiler efficiency, waste solids generation rate and disposal properties, erosion/corrosion potential, and turndown capabilities.

Coal is the fuel of major interest from an SO_2 emissions standpoint. Of the 80 existing or planned units in the U. S., coal is the sole design fuel in 14 units and is one of several design fuels in 9 units. Excluding boilers that are test, demonstration, undisclosed, or uncompleted units reduces this number to 8 commercially-operated, coal-fired AFBC units.

SO_2 Reduction Performance

Research on AFBC test units has shown that SO_2 reduction performance is dependent on many variables -- the most important include the Ca/S molar feed ratio, sorbent particle size and reactivity, and gas-phase bed residence time.

The SO_2 reduction capabilities that have been demonstrated by AFBC boilers in the industrial size category are summarized below:

- TVA conventional FBC boiler: 87 to 98 percent SO_2 removal at a Ca/S ratio of 3.0 and solids recycle ratios ranging from 0 to 1.5. This unit is a utility type design, however, with a higher freeboard than typical industrial boiler designs. The results may not be directly applicable to the industrial setting. Performance results are based on continuous emission monitoring (CEM) data collected over two periods of 12 and 15 hours duration;

- Georgetown University conventional FBC boiler: 85 percent SO_2 removal with Ca/S ratios of 3 to 6 and solids recycle ratios near 2. This performance is a conservative indication of FBC capabilities since the unit was operating under significant design/operational anomalies. SO_2 CEM data were collected over a 30-day test period;

- United Shoe Manufacturing Corporation two-stage FBC boiler: 90 percent SO_2 removal at a Ca/S ratio of 3.0. Performance was measured by EPA Reference Method 6 over a 3 hour period;
- Iowa Beef Processors staged bed FBC boiler: 82 percent SO_2 removal was achieved at a Ca/S ratio of 3.0. Steady-state operation of the FBC unit was not achieved during the tests. Performance was measured by EPA Reference Method 6 over a 9 hour period.
- West German circulating FBC boiler: 90 percent SO_2 removal at a Ca/S ratio of 3.0. Test method and duration were not specified;
- South Texas circulating FBC boiler: 95 percent SO_2 removal at a Ca/S ratio of 4.5 achieved on an FBC unit which is based on a conservative design. Test method and duration were not specified;
- Plant A circulating FBC boiler: 90 percent SO_2 removal at a Ca/S ratio of 3.5. Test method and duration were not specified.

NO_x and PM Reduction Performance

FBC boilers have demonstrated inherently low NO_x emissions relative to conventional boilers due to FBC's lower bed temperatures. For those industrial units for which data are available, FBC NO_x emissions have been consistently below 0.5 $lb/10^6$ Btu. Staged-beds and circulating FBC boilers appear to have the greatest potential for reducing NO_x emissions below this level. However, the major emphasis in FBC research to date has been on optimizing combustion efficiency and SO_2 control. Existing NO_x emission data do not represent long-term testing at conditions designed to produce low NO_x emissions. Although the exact mechanism is not currently understood, test unit data indicate a definite tradeoff between SO_2 and NO_x emission control for the use of staged combustion air. The interactions between SO_2 and NO_x must be further defined to establish optimum overall performance.

PM control on FBC boilers has been effected by cyclones followed by either a fabric filter or an electrostatic precipitator (ESP). Fabric filters have been used more widely for commercial applications than ESPs due to the low resistivity of entrained solids from FBC boilers. PM emissions of less than 0.05 $lb/10^6$ Btu have been routinely achieved with fabric filters.

FBC Algorithm

A cost algorithm has been developed for estimating capital and annual costs for conventional FBC systems over a wide range of boiler sizes and operating conditions. The bases of the algorithm are the FBC system designs and vendor-supplied cost estimates reported in the FBC ITAR. A comparison of the ITAR design with current operating system parameters shows that the design is representative of AFBC systems being offered commercially to industrial plant owners. Two-stage and circulating FBC designs were not considered due to the lower market penetration expected for these systems in the next five years. This is due primarily to the conservative nature of the industrial boiler market and the fact that these two designs are in an earlier commercialization stage than the conventional bubbling bed design.

The Westinghouse model for SO_2 capture by limestone in a fluidized bed has been used to project required Ca/S ratios as a function of SO_2 removal efficiency, limestone particle size and reactivity, and coal type. The Westinghouse model is felt to be the best instrument for projecting required Ca/S ratios as a function of SO_2 removal efficiency over the studied range of coal types and industrial FBC operating conditions. The model adequately accounts for sulfur capture by coal-ash alkali species and is in reasonable agreement with performance data from large operating systems.

Cost Comparisons Among Independent Estimates

The FBC algorithm design basis and methodology have been validated in part by comparison with independent estimates developed by five other organizations, four of which currently offer industrial-size FBC boilers on a commercial basis. Annual cost comparisons among the FBC algorithm projections and the three available estimates show very good agreement. All five vendor capital cost estimates are in agreement with the algorithm projections. This comparison of five independent estimates with the FBC algorithm projections lends added validity to the algorithm as a cost estimating tool.

Economic Competitiveness of FBC

FBC boiler system costs have been compared with costs for a conventional boiler equipped with an FGD system (i.e., lime spray drying) and with costs for a conventional boiler using low sulfur compliance coal. FBC costs are estimated with the cost algorithm described above. Lime spray drying has been chosen as the FGD technology over wet scrubbing systems because (1) the technology is being widely applied for SO_2 control among industrial boilers; (2) spray drying costs are representative of costs for wet FGD technologies throughout the studied size range; and (3) the technology is similar to FBC technology in its use of a calcium sorbent and production of a dry waste product. Costs for the competing SO_2 control options are estimated with analogous model boiler cost algorithms. Model boiler sizes of 50, 100, 150, 250, and 400 million Btu/hr are examined as representative of boilers operating in the industrial sector.

The purpose of these comparisons is to identify trends related to the relative competitiveness of the three options as SO_2 emission levels become more stringent. The absolute accuracy of individual capital and annual cost estimates is approximately ± 30 percent in keeping with the bases and methodology of the cost-estimating procedures. The accuracy of annual cost comparisons between technologies is less (near 15 percent) due to common operating and maintenance (O&M) cost items. Cost differences are felt to be significant if they exceed these limits. These cost differences are also dependent on the technical and economic assumptions that form their basis and thus should be used with caution in view of this and the overall accuracy level.

The SO_2 emission levels chosen for examination are 1.7, 1.2, and 0.8 lb $SO_2/10^6$ Btu. In addition, FBC and FGD options have been compared at SO_2 removal efficiencies of 65, 75, 80, and 90 percent. Removal efficiency levels for FBC and FGD are specified on the basis of the target emission level and coal fuel properties; compliance coals are selected to meet the emission levels (assuming continuous SO_2 monitoring) without the use of SO_2 control equipment. Emission levels for NO_x and PM are consistent for all SO_2 control altenratives examined.

The economic analysis results show that FBC system annual costs are not significantly different from those for the conventional boiler/FGD system (the FGD option) and compliance coal combustion (the CC option) for all boiler sizes and SO_2 emission levels examined. The annual cost differences between options do not exceed 15 percent, which is within the overall accuracy of the annual cost estimates. Capital costs for the three SO_2 control options were also comparable in all but the single case of a 50 million Btu/hr boiler operating to meet a 1.7 lb $SO_2/10^6$ Btu limit; capital costs for the CC option in this instance are significantly (i.e., greater than 30 percent) lower than the FBC option.

Comparing FBC and FGD system costs as a function of SO_2 emission limits, the results show that FBC competitiveness relative to FGD remains nearly constant as the SO_2 limitation becomes stricter for all boiler sizes based on the use of conservative Ca/S ratios. For optimistic Ca/S ratios, FBC competitiveness increases slightly with more stringent emission limits. This trend highlights the greater R&D incentives for lowering Ca/S ratios which will develop if SO_2 emission limits are reduced. Within a given emissions limit category, FBC competitiveness generally increases relative to FGD as boiler size decreases.

When comparing FBC with CC options the same general trends apply: (1) the relative cost competitiveness between the two alternatives remains nearly constant over the studied range of SO_2 emission limits and (2) FBC cost competitiveness decreases slightly as boiler size increases. Unlike the FBC-FGD cost comparison, however, FBC competitivenss relative to CC does not change significantly if Ca/S ratios are reduced to optimistic levels.

A second type of emission limit which currently applies to utility boilers with heat input capacities greater than 250 million Btu/hr is a requirement for a specific level of SO_2 removal. When FBC and FGD annual costs are compared at equal SO_2 reduction efficiencies between 65 and 90 percent, the results follow the same trend identified above: FBC competitiveness vis-a-vis FGD remains relatively unchanged over the studied range of SO_2 percentage removal requirements. If the optimistic Ca/S ratios are used for the FBC alternatives, FBC competitiveness increases as SO_2 removal levels become more stringent.

The conclusions drawn from these trends are that (1) studied cost differences between FBC technology, conventional boiler/FGD systems, and compliance coal combustion are projected to be small for the studied range of SO_2 emission limits and (2) that cost competitiveness among these technologies is not expected to change significantly as the emission limitations change. Absolute economic competitiveness among these options will be sensitive to site-specific parameters and decided on a case-by-case basis. Given the small cost differences among SO_2 control options, it is unlikely that economics alone will be the deciding factor when a choice is made. Rather less tangible factors such as operator requirements for fuel flexibility and preference for risk are likely to play a major role in the decision process.

To be significantly favored over competing SO_2 control options, the algorithm costing analysis indicates that FBC systems should be approximately 15 percent less expensive on an annual cost basis. This advantage over FGD systems could only be achieved by a reduction of FBC capital costs by about 50 percent relative to FGD for the case of a 150 million Btu/hr boiler operating to meet a 0.8 lb $SO_2/10^6$ Btu limit. reducing the FBC Ca/S ratio to a theoretical low of 1.0 would not be sufficient to account for this 15 percent differential. To achieve the same competitive edge over compliance coal combustion, low sulfur coals prices would have to rise almost 65 percent relative to high sulfur coal, or FBC relative capital costs would have to decline by over 60 percent, or a combination of the two shifts would have to occur. The likelihood of cost changes of this magnitude occurring in the foreseeable future as a result of coal market or technological changes is quite remote. As indicated, these changes apply to the case of a 150 million Btu/hr boiler and a 0.8 lb $SO_2/10^6$ Btu limit. Relative changes of a similar magnitude would be required for other boiler sizes and emission limits.

The coal price sensitivity of annual costs for the three SO_2 control alternatives are equivalent for practical purposes. For a 150 million Btu/hr boiler operated to meet a 1.2 lb $SO_2/10^6$ Btu emission limit, a $1.00/million Btu coal price increase will translate to an annual cost increase of approximately $800,000 for each technology, or about 13 percent.

SECTION 3

AFBC TECHNOLOGY STATUS

Atmospheric fluidized bed combustion technology (AFBC)has developed rapidly over the last four years. This section will focus on the technology developments concerning new bed configurations and improvement of emissions control, especially SO_2 and NO_x emissions. The advances which have resulted from both governmental and private research and development programs will be reviewed. A summary of the manufacturers offering AFBC units and existing and planned AFBC units will be presented. Finally, recent improvements in technology and projected technology trends related to SO_2 control, NO_x control, particulate control, solid waste disposal/utilization, and boiler performance will be discussed.

3.1 MECHANISMS FOR SO_2, NO_x, AND PM CONTROL

The Interagency Technology Assessment Report (ITAR) on fluidized bed combustion described the basic technology and pollution control capabilities of first generation AFBC boilers.[1] This section briefly reviews the information in the ITAR and updates it with recent developments related to boiler design and control of SO_2, NO_x, and particulates.

3.1.1 AFBC System Description

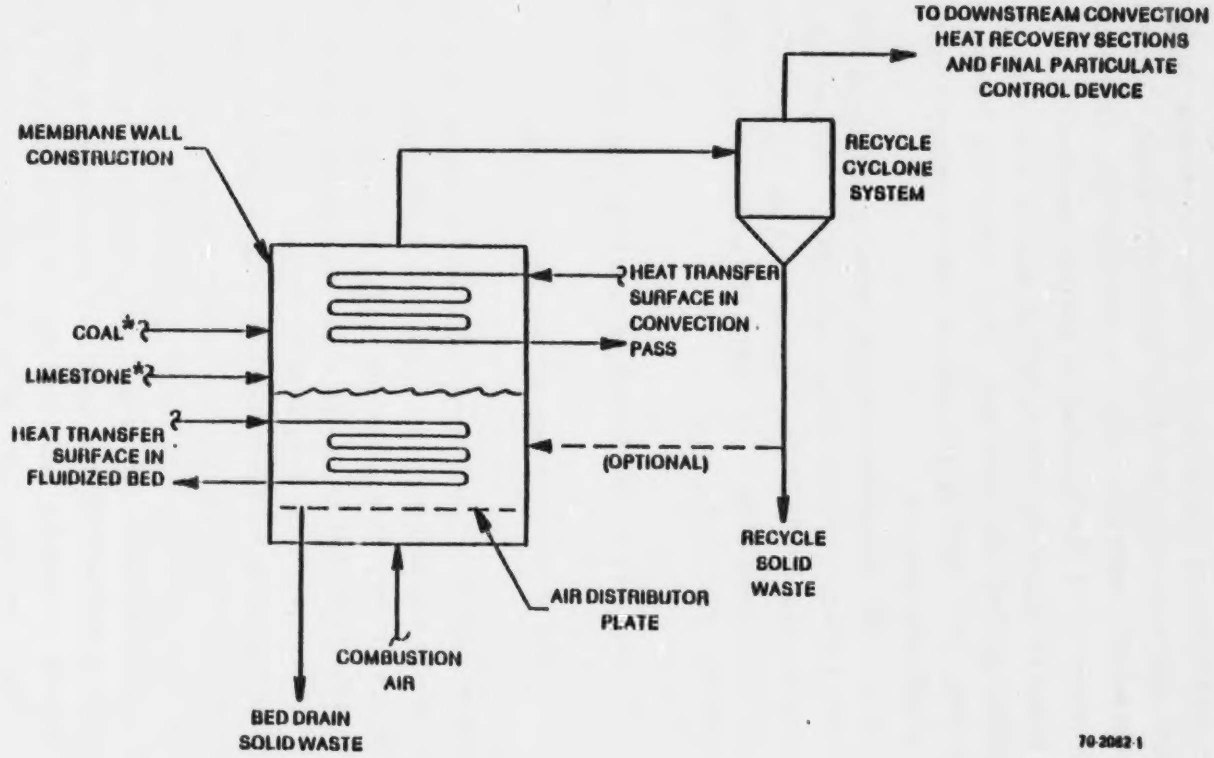
Atmospheric pressure fluidized bed combustion boilers are now commercially available in several different configurations. First generation units were based on a stationary bubbling bed design. Since the ITAR was published, a significant amount of development work has been conducted to more thoroughly investigate the beneficial effects of recycling elutriated bed material. Different configurations of AFBC boilers have become available as a result of recent changes in the fluidized bed design and/or the approach for utilization of the material removed from the flue gas. Design alternatives which have recently been implemented or are available on a commercial scale include the conventional bubbling bed with recycle, staging of combustion air, staged fluidized beds, and circulating fluidized beds. Pressurized fluidized bed combustion (PFBC) has been under development for several years, but has not yet been used in commercial applications. Therefore, the following discussion will focus on AFBC technology.

In the conventional bubbling bed system presented in Figure 3.1-1, fuel and sorbent, usually coal and limestone, are continuously fed into a bed of fluidized particles. The limestone is added for SO_2 removal. The fluidized bed consisting of unreacted, calcined, and sulfated limestone particles, coal, and ash is suspended in a stream of combustion air blowing upwards from an air distribution plate. Bed material is drained from the bed to maintain the desired bed depth. Some bed material is also elutriated from the bed with the combustion gas. This entrained material is separated from the flue gas by cyclones and a baghouse or electrostatic precipitator. The material is then discarded as a solid waste. A more detailed description of the conventional bubbling bed AFBC boiler is presented in the ITAR.[1]

In an AFBC boiler with solids recycle, flue gas with entrained bed material is passed through a primary cyclone where 80 to 90 percent of the entrained material is removed. All or part of this material is then fed back to the fluidized bed. The net effect of solids recycle is an increased fuel and sorbent residence time in the bed, with improvements in combustion efficiency and SO_2 and NO_x control.[2,3,4,5]

Staging of combustion air is a recently developed option which reduces NO_x emissions. A substoichiometric amount of air is added at the fluidizing air (primary air) injection point. The balance of the air needed to achieve adequate combustion efficiency is added above the bed. This allows combustion to be completed in the freeboard (i.e., space between the top of the fluidized bed and the boiler outlet). Early testing with staged combustion air showed NO_x reductions of up to 50 percent.[6] Testing has also shown, however, that an increase in SO_2 emissions occurs with staged combustion.[7,8]. (Refer to Section 4.3.)

A more complicated approach to isolate competing mechanisms in the fluidized bed is to actually operate the AFBC unit with two separate


TO DOWNSTREAM CONVECTION HEAT RECOVERY SECTIONS AND FINAL PARTICULATE CONTROL DEVICE
RECYCLE CYCLONE SYSTEM
MEMBRANE WALL CONSTRUCTION
HEAT TRANSFER SURFACE IN CONVECTION PASS
COAL*
LIMESTONE*
HEAT TRANSFER SURFACE IN FLUIDIZED BED
(OPTIONAL)
RECYCLE SOLID WASTE
AIR DISTRIBUTOR PLATE
COMBUSTION AIR
BED DRAIN SOLID WASTE
70-2062-1

*Coal and limestone may be fed above, in, or under the fluidized bed.

Figure 3.1-1. Conventional AFBC boiler flowsheet

fluidized beds. In this arrangement, one bed is stacked on top of the other. The lower bed is fed only coal and is operated at substoichiometric air conditions to reduce NO_x formation. Limestone is fed to the upper bed where desulfurization and final combustion occur. Since combustion and SO_2 retention/NO_x reduction occur in separate beds, conditions can be varied independently in the two beds to achieve the desired performance. Also, the distribution plate for the upper bed acts as a baffle, reducing fines elutriation from the lower bed. This lowers the freeboard requirements for both beds.[9,10] Baffles can also be used to reduce freeboard requirements for single bed boilers.

One of the more promising and recently developed AFBC technologies involves a circulating fluidized bed (CFB). Similar technology was originally used in other applications such as fluidized catalytic cracking of petroleum feedstocks. Two basic differences exist between CFB and conventional AFBC technology:

- the size of the limestone particles fed to the system, and
- the velocity of the fluidizing air stream.

Limestone feed to a conventional AFBC boiler ranges from fine particles (~500 μm) to coarse particles (~2000 μm). CFB technology is characterized by the use of very fine limestone particles (~200 μm and less). The conventional AFBC boiler design also incorporates relatively low superficial air velocities, ranging from 4 to 12 ft/sec. This creates a stable fluidized bed of solid particles with a well-defined upper surface. CFB technology, by contrast, employs superficial velocities typically ranging from 20 to 40 ft/sec. As a result, a physically well-defined bed is not formed; instead, solid particles (coal, limestone, ash, sulfated limestone, etc.) are entrained with the transport air/combustion gases. The solids are continuously circulated back into the combustion region, where fresh coal and limestone are fed. Simultaneously, solids are continuously removed from the system. CFB boiler systems are characterized by very high recirculated solids flow rates, up to three orders of magnitude higher than the combined coal/limestone feed rate.[11]

Many CFB boiler systems have been developed. Three representative systems, ranging in level of complexity, are discussed below.

The Pyropower design for industrial applications shown in Figure 3.1-2 features a combustion chamber of membrane wall construction and a refractory-lined hot cyclone collector.[12,13,14] The designer claims that a 3:1 turndown can be achieved by varying the air and fuel feed rates. Combustion chamber temperature is 1550°F. The circulation of solids allows for improved combustion efficiency and limestone utilization.

The Lurgi system shown in Figure 3.1-3 incorporates a separate fluidized bed economizer and evaporator for heat recovery.[11,15] Because much of the total heat recovery occurs in the cooler, turndown can be achieved by reducing the rate of solids circulation between the combustion chamber and the fluidized bed cooler.

The Battelle Multisolid Fluidized Bed Combustion (MS-FBC) process is depicted in Figure 3.1-4.[16,17,18] The process is characterized by a dense bed, an entrained bed, and a traditional fluidized bed. The stationary dense bed, located in the combustor, consists of an inert material with a relatively high specific gravity. These coarse particles are not entrained by the circulating gas, which has a velocity of 30 to 40 ft/sec. This bed serves to provide mixing of the coal/limestone feed with the combustion air and to contain the combustion zone. The entrained bed consists of fine particles of inert material that are continuously separated from the combustion gas and circulated back to the combustor. These fine particles accumulate in an external boiler as the third bed, a conventional fluidized bed operated at low superficial velocity, from 1 to 2 ft/sec. Little or no combustion occurs in the external boiler. Approximately two-thirds of the combustion heat energy is recovered by this external boiler. Additional flue gas energy is recovered in a downstream convection section. Turndown is achieved by reducing the flow of entrained bed material from the external boiler's fluidized bed to the combustor.

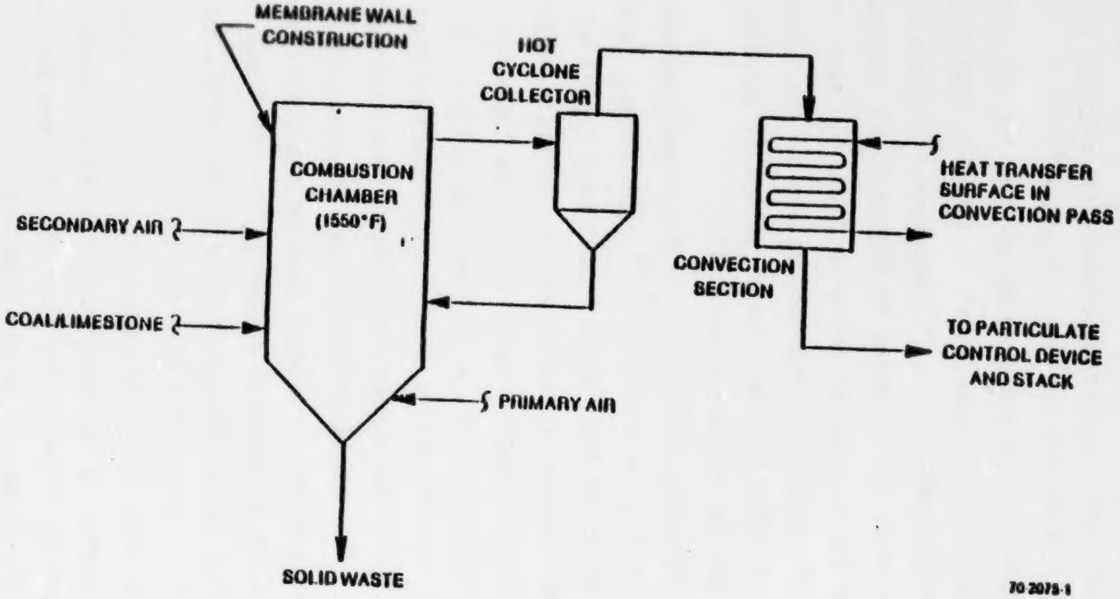

MEMBRANE WALL CONSTRUCTION
HOT CYCLONE COLLECTOR
COMBUSTION CHAMBER (1550°F)
SECONDARY AIR
COAL/LIMESTONE
PRIMARY AIR
SOLID WASTE
CONVECTION SECTION
HEAT TRANSFER SURFACE IN CONVECTION PASS
TO PARTICULATE CONTROL DEVICE AND STACK
70 2075-1

Figure 3.1-2. CFB boiler - Pyropower design (12,13,14).

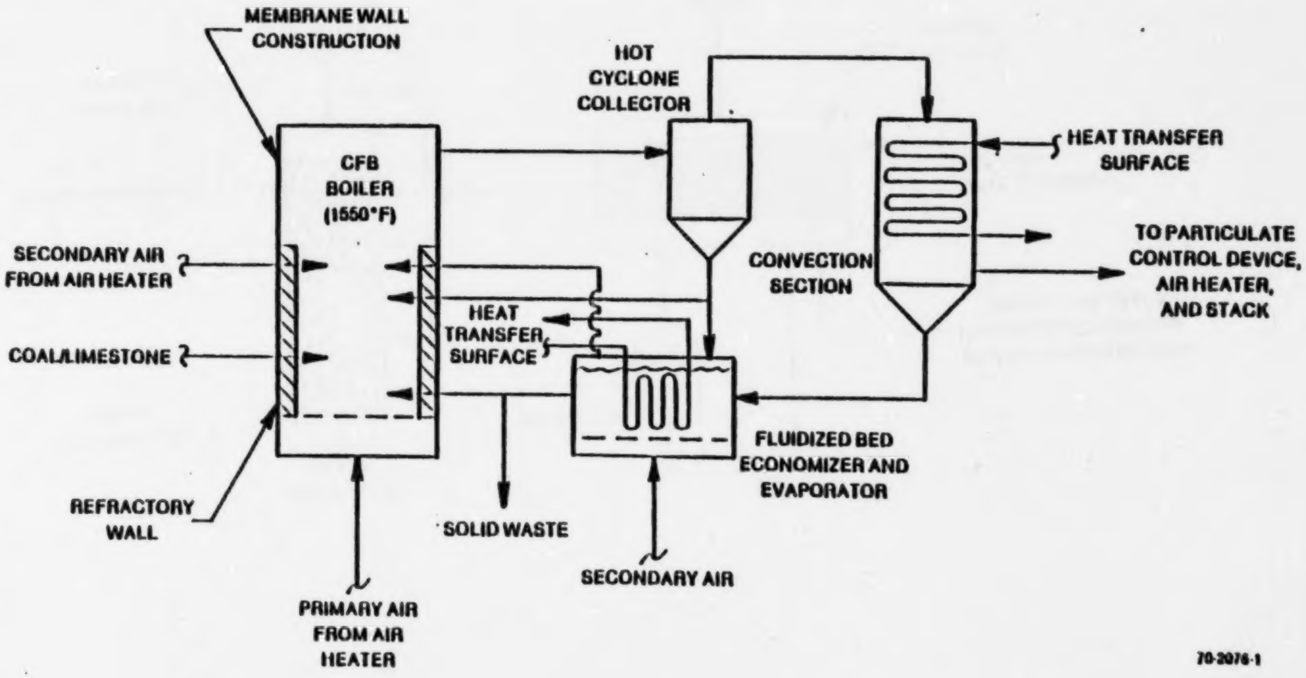

MEMBRANE WALL CONSTRUCTION
CFB BOILER (1550°F)
SECONDARY AIR FROM AIR HEATER
COAL/LIMESTONE
REFRACTORY WALL
PRIMARY AIR FROM AIR HEATER
HOT CYCLONE COLLECTOR
HEAT TRANSFER SURFACE
SOLID WASTE
CONVECTION SECTION
HEAT TRANSFER SURFACE
TO PARTICULATE CONTROL DEVICE, AIR HEATER, AND STACK
FLUIDIZED BED ECONOMIZER AND EVAPORATOR
SECONDARY AIR
70-2076-1

Figure 3.1-3. CFB boiler - Lurgi design (11,15)

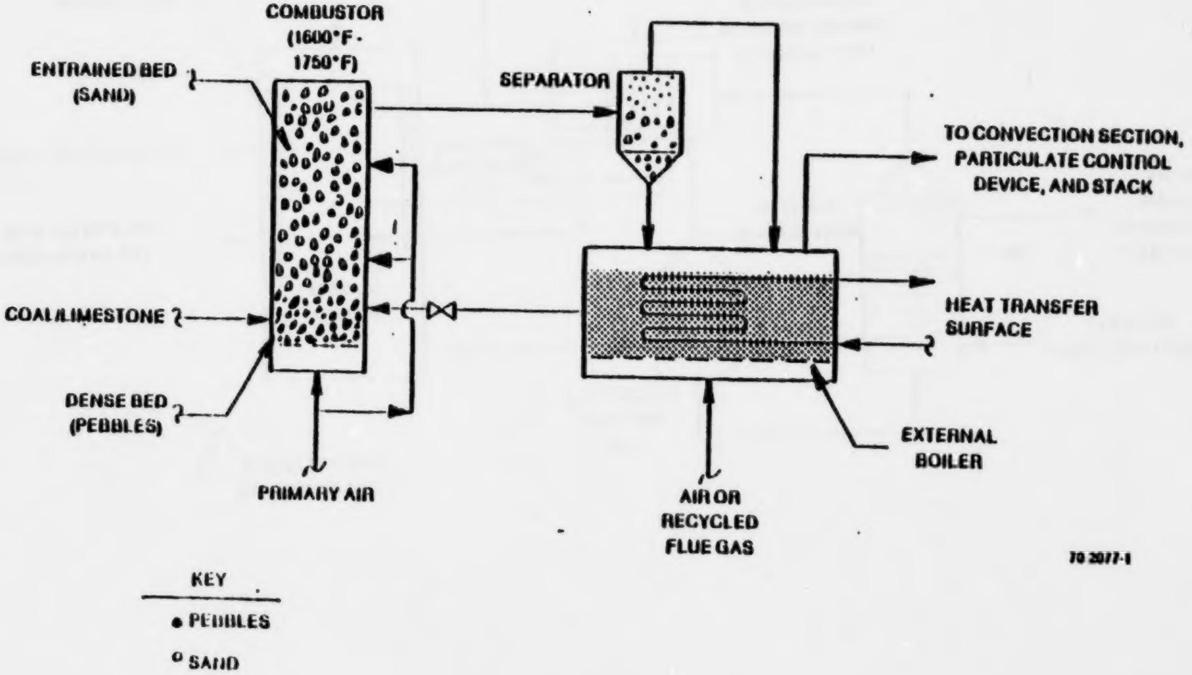

COMBUSTOR (1600°F - 1750°F)
ENTRAINED BED (SAND)
COAL/LIMESTONE
DENSE BED (PEBBLES)
PRIMARY AIR
SEPARATOR
TO CONVECTION SECTION, PARTICULATE CONTROL DEVICE, AND STACK
HEAT TRANSFER SURFACE
EXTERNAL BOILER
AIR OR RECYCLED FLUE GAS
70 2077-1
KEY
● PEBBLES
○ SAND

Figure 3.1.4. CFB boiler - Battelle's MS-FBC process (16,17,18)

Several advantages of the CFB process have been claimed over conventional AFBC technology:

- higher combustion efficiency, exceeding 99 percent;
- greater limestone utilization, due to recycle of unreacted sorbent and to the limestone feed size (greater than 85 percent SO_2 removal efficiency is projected with a Ca/S ratio of about 1.5, with the potential for greater than 95 percent SO_2 removal efficiency);[11,15,17,18]
- simple turndown and excellent load following capabilities;
- lower NO_x emissions because of staged combustion (less than 100 ppm NO_x are projected);[7,11,15]
- less critical coal feed design, since high velocities ensure good mixing;
- potentially fewer corrosion problems, since heat transfer surface is less likely to be located in reducing zones;
- minimal excess air requirements, since the high velocities promote good mixing and combustion efficiency;
- less dependence on limestone type, since reactivity is improved with the fine particle sizes; and
- reduced solid waste rates, because of lower limestone requirements.

Potential drawbacks of the technology include:

- increased capital costs;

- greater energy losses due to high pressure drops across the system;

- a combustor height of 30 to 100 feet;[11,15]

- uncertainty regarding the hot cyclone's ability to effect the required solids/gas separation and to resist erosion and corrosion; and

- erosion of components subjected to impingement of high velocity particles.

CFB technology has reached the commercialization stage, with several boilers now in operation in the U.S. and Europe. These boiler designs have been used for both retrofit and new installations. In this country, Battelle's MS-FBC process has been identified as having distinct advantages over conventional boiler technology for use in thermally enhanced oil recovery (TEOR) steam generation applications burning solid fuels. TEOR requires 80 percent quality steam at 2500 psia. Generally, water with high total dissolved solids (TDS) is used once-through to generate this steam. Steam in the outlet tubes of the steam generator occupies about 95 percent of the tube volume. Steam in conventional boiler outlet tubes may occupy only 18-20 percent of the volume due to the high recirculation ratio. The conditions of high steam volume in the outlet tubes and high TDS, once-through water can lead to dry wall conditions, solids deposition on the tube wall, and rapid tube burnout if average or point heat fluxes become excessive. Conventional drum type boilers were tried on TEOR projects and were removed because of operating difficulties and/or excessive operating costs due to rapid tube burnout and high quality feed water requirements. The decoupled external heat exchanger in the MS-FBC process utilizes fluidized bed heat transfer techniques to permit precise control of heat fluxes. In addition, the external heat exchanger allows the heat transfer to be controlled without affecting combustor performance.[19,20]

The recycle, staged, and circulating bed configurations have all been applied commercially in the past four years. In addition, the Department of Energy (DOE) is funding advanced FBC technologies that, if proven feasible, might substantially improve fluidized bed systems now on the market. Concepts such as ultra-high velocity combustors, staged cascades, or advanced circulating beds might well be the basis for the fluidized bed systems of the 1990's and beyond.[21]

One configuration that is receiving considerable development effort and DOE funding, but has not yet been commercialized, is pressurized fluidized bed combustion (PFBC). PFBC has the potential to have the lowest bus-bar energy cost of any near-term coal utilization option for electrical power generation.[23] In a PFBC boiler design, the combustion chamber operates at 5 to 20 atmospheres, with the cleaned exhaust gases driving a gas turbine. Potential advantages of the technology include:

- a smaller boiler, due to better heat transfer in the bed;
- lower sorbent feed rates, because the sulfation reaction is favored at high pressures; and
- increased cycle efficiency, especially when applied to a combined cycle as depicted in Figure 3.1-5.

Issues which have contributed to a lag in the commercial development of PFBC as compared to AFBC technology include (1) the ability of the flue gas cleanup device to reduce solids loadings to the gas turbine to acceptable levels, and (2) the increased complexity of the process.

3.1.2 Mechanisms for SO_2 Control

The ITAR identified the following factors as being important to the control of SO_2 emissions:

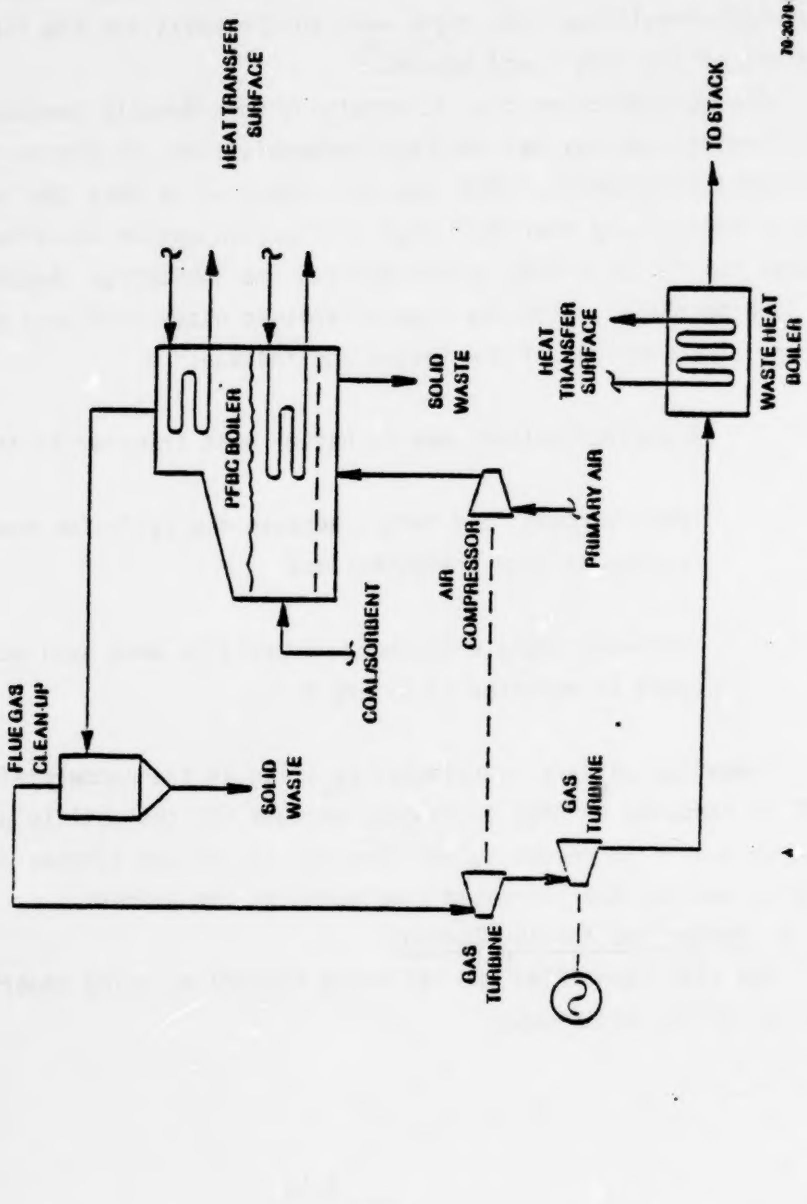

HEAT TRANSFER SURFACE
78-2070-1
TO STACK
PFBC BOILER
SOLID WASTE
HEAT TRANSFER SURFACE
WASTE HEAT BOILER
PRIMARY AIR
AIR COMPRESSOR
COAL/SORBENT
FLUE GAS CLEAN-UP
SOLID WASTE
GAS TURBINE
GAS TURBINE

Figure 3.1-5. PFBC direct-fired combined cycle

- Primary Factors
  - Ca/S molar feed ratio
  - sorbent particle size
  - gas phase residence time
- Secondary Factors
  - sorbent reactivity
  - bed temperature
  - feed mechanisms
  - excess air

Detailed information on the impact of these factors on SO_2 emissions can be found in the ITAR.[1]

These factors can be varied to optimize sulfur capture. However, it should be emphasized that these factors also affect other important performance variables, including boiler operation (e.g., combustion efficiency, boiler efficiency, etc.) and control of other emissions (e.g., NO_x, particulates, and solid waste). Therefore, a number of important design compromises must be made between boiler performance and environmental impact.

Recent designs have become more sophisticated in response to needs for optimizing the tradeoffs resulting from coupling combustion and in-situ emissions control. Recycle of elutriated material, staged combustion air, staged beds, and circulating beds affect SO_2 emissions and other performance variables.

Recycle of elutriated material improves SO_2 capture by providing longer limestone residence time in the bed, increasing limestone utilization. Longer residence time is also provided for unburned coal particles which improves combustion efficiency and tends to reduce NO_x emissions.

Staged combustion air reduces NO_x emissions. However, SO_2 emissions increase with staged combustion due to the creation of a reducing zone in the combustor which shortens the length of the oxidizing region. This limits the extent of the CaO-SO_2-O_2 to $CaSO_4$ reaction. A tradeoff between NO_x and SO_2 emissions results.[7,23] (Refer to Section 4.3.)

Staged beds decouple the design tradeoffs associated with a one-bed unit and allow combustion and emissions control to be optimized more independently.

The operating conditions present in circulating bed AFBC boilers differ from those in conventional bubbling bed AFBC boilers. The smaller limestone feed size promotes limestone utilization. Smaller limestone particles are sulfated to a greater degree than large particles, resulting in improved SO_2 retention for a given amount of limestone. The recycle of unreacted limestone and unburned coal increases SO_2 removal and combustion efficiency by increasing residence time. Recycle also permits attrition of the limestone particle which further enhances SO_2 absorption and limestone utilization. Higher superficial velocities result in turbulence and better mixing. This increases the contact between SO_2 and CaO as well as the contact between NO_x and carbon. Carbon reduces NO_x to N_2. Thus, lower emissions of SO_2 and NO_x are obtained. Staging of combustion air can also be used with the circulating bed design, but the tradeoff between SO_2 and NO_x performance still exists.[7,8] (Refer to Section 4.3.)

Another important point that should be discussed based on recent test data is the effect of coal characteristics on SO_2 emissions. In addition to the sulfur content, the form of the sulfur and the alkalinity and quantity of ash can affect SO_2 emissions. Tests conducted by DOE's Morgantown Energy Technology Center (METC) and Grand Forks Energy Technology Center (GFETC) on low-rank fuels indicate that some lignites and low-sulfur subbituminous western coals contain a significant quantity of calcium and sodium alkalinity in the ash.[24,25] The relatively large quantities of alkaline ash and low sulfur content combine to provide significant sulfur capture. The inherent SO_2 control reduces the amount of limestone that must be introduced to obtain high SO_2 removal efficiencies. In fact, 90 percent SO_2 removal can be achieved without any limestone addition.[25] However, it is also important to note that the overall heat release rate per ton of input materials for low-ranked coals is about equal to that for higher quality coals with limestone addition. A design tradeoff that must be considered is the increasing agglomerating tendencies of the fuels containing high sodium levels. The sodium combines with silica and other elements to form low-melting temperature ash. The ash particles become soft and agglomerate into larger particles. Agglomeration can eventually result in loss of fluidization at some operating conditions. Agglomeration can be minimized by several methods, including bed flushing, lowering operating temperatures, raising gas velocities, operating without recycle, and adding alkali suppressants.[25]

3.1.3 Mechanisms for NO_x Control

The formation and control of NO_x in AFBC units is influenced by the following design factors, as mentioned in the ITAR:

- bed temperature,
- excess air,
- gas residence time,
- fuel nitrogen,
- coal particle size,
- superficial gas velocity, and
- bed composition (Ca/S ratio).

Although each of the operating parameters discussed above affects NO_x emissions, the primary goals of high combustion efficiency and SO_2 capture rather than low NO_x emissions tend to determine operating conditions.

Low NO_x emissions have been demonstrated for AFBC units in various studies, but the majority of the research work has been concerned with SO_2 emissions and combustion efficiency. The optimization of parameters affecting SO_2 emissions and combustion efficiency does not necessarily reflect optimum conditions for the reduction of NO_x emissions. Recent test data, especially for some of the new design configurations, demonstrate the capability of AFBC units to achieve low NO_x emissions. These data will be discussed in Section 4.

The ITAR discussed the fact that the lower combustion temperature in AFBC boilers (1400° to 1650°F) as compared to stoker and pulverized coal combustion boilers (greater than 2000°F) reduces the level of NO_x emissions. Most of the NO_x formed in AFBC units is due to the oxidation of fuel nitrogen; the rate of formation of thermally fixed NO_x from combustion air is very slow due to the low combustion temperature. More recent research has suggested that NO_x formation in fluidized bed combustors is due primarily to oxidation of non-volatilized nitrogen-containing compounds in the char.[26] Other research suggests that it is both non-volatile and volatile nitrogen compounds which contribute to NO_x formation.[27,28]

Several researchers have shown that the initial NO_x concentration in an AFBC bed rises rapidly as flue gas moves upward from the point of air/fuel injection.[27,29,30] The NO_x concentration then decays at the top of the bed and in the freeboard area, indicating that NO_x is reduced by reaction with other species present.

The reactions of NO with carbon at temperatures above 1400°F apparently contribute to this phenomenon. These reactions are of the following forms:

$$C + NO \rightarrow \tfrac{1}{2}N_2 + CO \qquad (3\text{-}1)$$

$$CO + NO \rightarrow CO_2 + \tfrac{1}{2}N_2 \qquad (3\text{-}2)$$

Below bed temperatures of 1450° to 1500°F, homogeneous reactions between gas phase carbon (i.e., carbon monoxide) and NO_x are thought to predominate. Above 1500°F, heterogeneous reactions between gas phase NO_x and solid phase carbon in char particles are thought to be the predominant mechanism for NO_x reduction.

Some investigators have found evidence that the reduction of NO by CO (Eq. 3-1) may be catalyzed by the presence of $CaSO_4$ in the bed.[6,31] Also, calcium compounds may take part directly as a reactant, by the following reaction scheme:

$$CaO + SO_2 \rightarrow CaSO_3 \qquad (3\text{-}3)$$

$$2CaSO_3 + 2NO \rightarrow 2CaSO_4 + N_2 \qquad (3\text{-}4)$$

As previously mentioned, the more recent sophisticated design configurations provide advantages for NO_x control as well as SO_2 control and combustion efficiency. Recycle of elutriated solids has replaced the carbon burnup cell as a means to increase combustion efficiency. Carbon in the recycled char is available for heterogeneous reduction reactions between NO_x and char.[2,3,5,32] Increased freeboard heights provide greater contact time to promote NO_x reduction reactions. Staged beds allow conditions in the two beds to be varied independently to reduce NO_x emissions. Operation of the lower bed at sub-stoichiometric air rates reduces NO_x formation; char in the upper bed enhances the rate of NO_x reduction reactions. Circulating bed AFBC units feature extensive recirculation of elutriated solids and staging of combustion air which serve to lower NO_x emissions, as previously stated. Staging the combustion air in a conventional bubbling bed AFBC promotes heterogeneous and homogeneous reduction of NO_x in the fuel-rich bed.

3.1.4 Mechanisms for Particulate Control

Both fabric filters and ESPs have been considered for final particulate matter control after primary control of entrained solids with one or more cyclones. The majority of AFBC units in existence utilize fabric filters. The low resistivity of AFBC ash and calcium solids and the fluctuating operating conditions, especially during startup and turndown, limit the effectiveness of ESPs. Only limited research on PM control has been conducted in the past since fabric filters have proven to be effective. However, the Tennessee Valley Authority/Electric Power Research Institute (TVA/EPRI) 20 MWe pilot plant will test ESP performance in the future using a small slip stream of flue gas.[33]

3.2 STATUS OF DEVELOPMENT

This section deals with the status of AFBC with respect to research and development and projected technology trends. Manufacturers currently offering commercial AFBC units, along with existing and planned units, are presented.

3.2.1 U.S. DOE Development Programs

The U.S. Department of Energy (DOE) is sponsoring AFBC research at the facilities listed in Table 3.2-1. The areas of research for each facility are also provided in the table. The research in the pilot programs is generally directed at the fundamental properties, rates, and mechanisms of AFBC systems as well as testing the feasibility of using low-grade fuels and alternate sorbents. DOE demonstration programs have taken place at the sites listed in Table 3.2-2. These programs were designed to prove the commercial feasibility of AFBC technology and its ability to burn different types of coal in an environmentally acceptable manner. Since commercial feasibility has been shown, DOE is leaving the commercial development of existing technology to private industry and is now initiating research investigating novel FBC methods considered too risky for private industry to undertake.

Pressurized fluidized bed combustion (PFBC) is an example of a new technology for which DOE is sponsoring research. DOE-sponsored studies on PFBC are taking place at the IEA Grimethorpe Facility and the Coal Utilization Research Laboratory in England, at the General Electric LTMT Facility in New York, and at New York University. More information on these PFBC facilities, along with private PFBC research facilities, is listed in Table 3.2-3. PFBC boilers have the potential for combined cycle generation

TABLE 3.2-1. SUMMARY OF DOE PILOT PROGRAMS[44]

Facility	Location	Diameter, Inches	Research Emphasis
Morgantown Energy Technology Center	Morgantown, W. Va.	4 6 18	An extensive program of low-grade fuel studies, which includes anthracite refuse, high-sulfur coals, lignites, oil shales, and discarded tires, is in progress to provide operational design data and demonstrations of low-grade fuel feasibility.
Brookhaven National Laboratory	Long Island, NY	1,6	Activity is aimed at developing an SO_2 sorbent, using commercial silicate-bearing portland cement for desulfurizing FBC gases. Once through, as well as regenerative, systems are being evaluated. Basic data on the kinetics and mechanisms of the reactions occurring in the combustor and regenerator are obtained as required.
Argonne National Laboratory	Argonne, IL	6	Projects provide basic support information for FBC development in the general areas of improved combustion efficiency, NO_x emission control, and limestone utilization.
Oak Ridge National Laboratory	Oak Ridge, TN	10	Data concerning elutriated char utilization are being gathered and processed.

TABLE 3.2-2. SUMMARY OF DOE DEMONSTRATION PROGRAMS

Facility	Objectives	Size	Emission Controls	Distinguishing Characteristics
Georgetown University - Washington, D.C. - Vendor/A&E--Foster Wheeler Energy Corp./Pope, Evans, and Robbins - Startup--July 1979 - Still operating	Demonstrate industrial and institutional application of FBC using high sulfur coal in an acceptable manner in a populated area.	100,000 lb/hr of steam 2-106 ft^2 bed area ~$110x10^6$ Btu/hr	Limestone addition for sulfur capture (Ca/S = 3 to 6) Solids Recycle Baghouse for PM control	Stoker overbed coal feed, above-bed gravity limestone feed 1550°F bed temperature 8 ft/sec gas velocity 4.5 ft bed depth Operated for 1600 hrs. in compliance with D.C. regulations
Alexandria Pilot Development Unit - Alexandria, Va. - A&E--Pope, Evans, and Robbins	Provide original design for Rivesville unit (listed below).	800 lb/hr coal 3 ft x 3 ft bed 0.5 MW ~$10x10^6$ Btu/hr	Limestone addition for sulfur capture (Ca/S = 3) Solids Recycle Baghouse for PM control	5 to 12 ft/sec gas velocity Tests conducted using different fuels
U.S. Navy Great Lakes Training Facility - Great Lakes, Ill. - Built by C-E Power Systems - Startup--September 1981 - Still operating	Demonstrate practicality of industrial FBC for high sulfur Illinois coal in an environmentally acceptable manner and appraise performance, reliability, and economics.	50,000 lb/hr of steam 140 ft^2 bed area $70x10^6$ Btu/hr	Limestone addition for sulfur capture (90 percent with Ca/S = 2.2, and 98 percent with Ca/S = 4 in subscale tests) Solids Recycle Baghouse for PM control	7 ft/sec gas velocity 3 ft bed height 1550°F bed temperature
Rivesville Unit - Rivesville, W. Va. - Built by Foster-Wheeler/Pope, Evans, and Robbins - Startup -- September 1976 - Dismantled 1980	Initial design of a multicell boiler to be used as a basis for a larger demonstration and utility-scale plant.	300,000 lb/hr of steam Total bed size: 460 ft^2 ~$450x10^6$ Btu/hr	Limestone addition for sulfur capture (Ca/S = 3-5) Solids Recycle Cyclones and electrostatic precipitator for PM control	Test plan concluded Four cells
Shamokin Area Industrial Corp. - Shamokin, Pa. - Built by E. Keller/Dorr-Oliver - Startup--August 1981 - Still operating	Test feasibility of using anthracite culm over wid range of operating conditions while satisfying air pollution control requirements	23,000 lb/hr of steam 100 ft^2 bed area ~$28x10^6$ Btu/hr	Limestone addition for sulfur capture Solids Recycle Cyclones and baghouse for PM control	3.5 ft/sec to 5.5 ft/sec gas velocity 3 ft to 5 ft bed height 1450°F to 1650°F bed temperature
East Stroudsberg State College - East Stroudsberg, Pa. - Built by Fluidyne Engineering Corp./International Boiler Works - Still operating	Scale-up of Shamokin unit. Demonstrate feasibility of using anthracite culm as fuel.	40,000 lb/hr ~$48x10^6$ Btu/hr		Anthracite culm fuel.
City of Wilkes-Barre - Wilkes-Barre, Pa. - Still operating	Scale-up of Shamokin unit. Demonstrate feasibility of using anthracite culm as fuel.	60,000 lb/hr ~$72x10^6$ lb/hr		Anthracite culm fuel.

TABLE 3.2-3. PFBC RESEARCH FACILITIES IN EXISTENCE OR UNDER CONSTRUCTION[34]

Organization	Argonne National Laboratory	New York University	Exxon Research and Engineering	NASA Lewis Research Center	Coal Utilization Research Lab (CURL)	Coal Utilization Research Lab (CURL)	GE LMT Facility
Location	Argonne, IL	Westbury, NY	Linden, NJ	Cleveland, OH	Leatherhead, England	Leatherhead, England	Malta, NY
Thermal Rating, (MWt)	0.15	7	1.7	0.5	0.2	6	0.45
Status	Operational 1982[a]	Operational 1983	Decommissioned	Decommissioned	Operational	Operating	Operational 1982
Operating and Design Parameters:							
Bed Plan Sect. (ft)	0.5 Dia.	2.5 Dia.	1.05 Dia.	0.75 Dia. for 3 ft Taper to 1.7 (top 7 ft)	1.0 Dia.	2 x 3 or 4	1.0 Dia.
Bed Plan Area (ft^2)	0.2	4.9	0.8	0.44 to 2.3	0.8	6 or 8	0.8
Expanded Bed Depth (ft)	3	12	10-14	2-8		12	5.3
Air Flow (lb/s)	0.25	4.0	1.0	0.17		2.0-4.0	0.44
Max. Shell Pressure (psia)	165	.50	147	120	75	88	150
Max. Bed Temperature (°F)	1800	1750	1800	1600		1750	1750
Max. Fluidizing Velocity (ft/s)	6	8	7	7		7	3
Coal Feed (lb/h)	20	2000	300	80	50	1700	131
Steam Temperature (°F)	--	(Water or Air)	(Water)	(Water)		(Water)	(Water)
Steam Pressure (psia)		(Water or Air)	(Water)	(Water)		(Water)	(Water)
Clean-up Equipment	3 Cyclones + Metal Filter	2 Cyclones + Baghouse	3 Cyclone Stages	2-in-1 Cyclone		Up to 3 Cyclone Stages	3 Cyclone Stages

List of Equivalents: 1 ft = 30.5 cm; 1 ft^2 = 0.0929 m^2; 1 lb = 454 g (mass); 1 psi = 6.895 kPa; °C = 0.586 (°F-32); 1 lb/h = 0.454 kg/h

TABLE 3.2-3. PFBC RESEARCH FACILITIES IN EXISTENCE OR UNDER CONSTRUCTION[34] (Continued)

Organization	Technical University, Warsaw	Curtiss-Wright Corp.	Curtiss-Wright Corp.	University of Natal	Combustion Power Co.	International Energy Agency	American Electric, Power, STAL-Level, Deutsche Babcock
Location	Warsaw, Poland	Wood-Ridge, NJ	Wood-Ridge NJ	Durban, South Africa	Menlo Park, CA	Grimethorpe, England	Malmo, Sweden
Thermal Rating, (MWt)	3	40	2.3	2	8	85	15
Status	Operational 1981	Standby	Operational Standby	Under Construction	No Longer Burning Coal	Operational 1981	Operational 1982
Operating and Design Parameters:							
Bed Plan Sect. (ft)		12 Dia.	3 Dia.	1.64 Dia.	7 Dia.	6.5 x 6.5	
Bed Plan Area (ft^2)	3.2	113	7.1	2.1	39.4	42.9	20 (at top)
Expanded Bed Depth (ft)		16	16	5.6	2	10	13
Air Flow (lb/s)		40[a]	2.3[a]		22.7	68	
Max. Shell Pressure (psia)	96	100	95	105	55	175	235
Max. Bed Temperature (°F)		1650	1650		1550	1740	
Max. Fluidizing Velocity (ft/s)	10	2.7	2.7		6.7	8.2	
Coal Feed (lb/h)	1100	9000	585		2000	22,000	5000
Steam Temperature (°F)		(Air Cooling)	(Air Cooling)		(Adiabatic)	824	
Steam Pressure (psia)		(Air Cooling)	(Air Cooling)		(Adiabatic)	440	
Clean-up Equipment		Recycle Cyclone + 3 Cyclone Stages	Recycle Cyclone + 3 Cyclone Stages		2 Cyclone Stages	2 Cyclone Stages	3 Cyclone Stages

List of Equivalents: 1 ft = 30.5 cm; 1 ft^2 = 0.0929 m^2; 1 lb = 454 g (mass); 1 psi = 6.895 kPa; °C = 0.586 (°F-32); 1 lb/h = 0.454 kg/h

[a] Combustion air only; in addition, twice this amount flows through the cooling coils.

of electricity by expanding the cleaned flue gas in a turbine generator and by expanding the steam generated from flue gas heat recovery in a steam turbine.

3.2.2 Other Development Programs

Numerous AFBC research facilities are owned and operated by private industry in the U. S., as listed in Table 3.2-4. Foreign private and government research facilities are listed in Table 3.2-5. These facilities are capable of performing tests at a wide variety of operating conditions in configurations ranging from the conventional bubbling bed to the circulating bed. Research and development conducted by private industry is directed more at the optimization of parameters affecting AFBC operation.

Of notable significance are the research programs sponsored by the Tennessee Valley Authority and the Electric Power Research Institute. Even though these organizations are primarily concerned with utility application of AFBC systems, much of the data generated is useful for evaluating the performance of AFBC boilers for industrial applications. TVA and EPRI are currently performing tests on a 20 MWe pilot plant in preparation for scale-up to a 100-200 MWe demonstration plant. One of the major goals of the testing is to demonstrate the environmental control capability of the unit as a basis for evaluating the environmental acceptability of AFBC on a commercial basis.[33]

3.2.3 Commercial Availability of AFBC

Domestic AFBC boiler manufacturers, along with their equipment specifications, are listed in Table 3.2-6. Foreign AFBC manufacturers are listed in Appendix E. The domestic units offered range in size from 2,000 to 600,000 lb/hr of steam at pressures and temperatures of up to 2650 psig and 1050°F, respectively (2.3 to 935 x 10^6 Btu/hr). The configurations available include the conventional bubbling bed, with or without recycle, the fully circulating bed, and staged beds. They can be designed to burn either a single or multiple fuels. Retrofit units are also offered by a few of the manufacturers. Many vendors are offering guaranteed systems for a

TABLE 3.2-4. EXISTING PRIVATE AFBC RESEARCH FACILITIES-UNITED STATES

Owned By	Location	Cross-Section Feet	Maximum Feed Rate			Superficial Velocity ft/sec
			Heat Input[a] 10^6 Btu/hr	Coal, lb/hr	Sorbent, Ca/S	
Babcock & Wilcox	Alliance, OH	1x1	0.72	60	10	4 to 12
Babcock & Wilcox	Alliance, OH	3x3	6.0	500	10	4 to 12
Babcock & Wilcox	Alliance, OH	6x6	24.0	2000	10	4 to 12
Battelle	Columbus, OH	.5D[b]	0.48	40	20	20 to 40
Battelle	Columbus, OH	.75D	0.60	50	20	
Battelle	Columbus, OH	1.25x2	4.8	400	150	20 to 40
Battelle	Columbus, OH	2D	2.4	200	75	6 to 10
Combustion Engr.	Windsor, CT	2.5x2.75	3.4	280	85	6 to 12
Combustion Power	Menlo Park, CA	1.7D	0.72	60	20	6
Combustion Power	Menlo Park, CA	2.5D	2.2	180	60	6
Combustion Power	Menlo Park, CA	3.0D	5.0	420	150	6
Fluidyne Engr.	Minneapolis, MN	1.5x1.5	0.60	50	20	
Fluidyne Engr.	Minneapolis, MN	1.5x1.5	0.60	50	20	
Fluidyne Engr.	Minneapolis, MN	3.5x5.5	7.6	630	250	2.5 to 4
Foster Wheeler	Livingston, NJ	1.7x1.7	6.0	500	200	5 to 14
Garrett	Torrance, CA	2D	2.4	200	75	4 to 6
General Atomics		1.3x1.3				4 to 12
General Electric		2x2				8 to 20
Johnston Boiler	Ferrysburg, MI	5x7.5	14.4	1200	400	
Mass. Inst. Tech.	Cambridge, MA	2x2	1.8	150	50	
Tenn. Valley Auth.	Shawnee, KY	12x18	264	22000		4 to 12
Univ. North Dakota[c]	Grand Forks, ND	0.5D				
Univ. North Dakota	Grand Forks, ND	1.5D				
Univ. North Dakota	Grand Forks, ND	3D				
Virginia Poly. Inst.	Blacksburg, VA	1.5x3	9.6	800	400	

[a] Assumes coal heat content at 12,000 Btu/lb.

[b] D = Diameter.

[c] Formerly Grand Forks Energy Technology Center.

TABLE 3.2-5. EXISTING AFBC RESEARCH FACILITIES - FOREIGN

Owned By	Location	Cross Section ft.	Size, MW	Superficial Velocity, ft/sec
UK National Coal Board	Marden Herefordshire	5.0 ft D	2.3	7.5
UK National Coal Board	Bury, Lancashire	4.4 ft D	1.8	8.9
UK National Coal Board	Newcastle-under Lyme, Staffordshire	9.0 x 7.5	9.5	-
Wallsend Slipway Engineers Ltd.	Edmonton, North London	6.2D	3.8	8.9
UK National Coal Board	Cheltenham, Gloucestershire	6.2 x 6.2	5	8.2
Department of Energy Conversion	Goteberg, Sweden	10.4 x 10.4	15.7	8.2
TNO/Stork Boilers	Netherlands	2-3 x 3	4	3.3 to 9.8
Swedish Board for Energy Source Development	Sweden	2.3 x 2.3	2.5	24.7

TABLE 3.2-6. DOMESTIC AFBC MANUFACTURERS[35]

Company Address	AFBC Boiler Technology		Watertube or Firetube Boiler	Types of FBC Systems Offered	Boiler Capabilities Commercially Available					Number of Units Installed	
	Built Under License	Licensing Company			Heat Input,[1] x10^6 Btu/hr	Steam Capacity, x1000 lb/hr	Pressure psig	Temperature, °F.	Fuel(s)	USA	Total
Babcock & Wilcox Co 20 S. Van Buren Ave. Barberton, OH 44203	No	-	Wt	Fx	More than 78	More than 50	150-2400	Up to 1050	-[2,3]	2	2
C-E Natco 5330 E. 31st St. Tulsa, OK 74135	Yes	Energy Resources Co.	-[4]	-[4]	-[4]	-[4]	-[4]	-[4]	-[4]	1[5]	1[5]
C-E Power Systems 1000 Prospect Hill Rd. Windsor, CT 06095	No	-	Wt	Fx, Fcb	60-750	50-500	100-1800	330-950	-[2]	1	1
Curtiss-Wright Corp. One Passaic St. Wood-Ridge, NJ 07075	No	-	Wt	Fx	24-180	20-125	100-800	250-825	Coal Wood-waste Biomass	1[6]	1[6]
Dedert Corp. Thermal Processes Div. 20000 Governors Dr. Olympia Fields, IL 60461	No	-	Wt, Ft	Fx	6-180	5-125	10-900	212-825	-[2]	2	2
Dorr-Oliver Inc. 77 Havemeyer Lane Stanford, CT 06904	No	-	Wt	Fx, Pcb	Up to 350	40-250	Up to 800	Up to 750	-[2]	1[7]	1[7]
Energy Products of Idaho 4006 Industrial Ave. Coeur d'Alene, ID 83814	No	-	Wt, Ft	Fx	12-380	10-250	15-1000	250-900	-[2]	18	22
Energy Resources Co. One Alewife Place Cambridge, MA 02140	No	-	Wt, Ft	Fx	Up to 360	10-250	Up to 1500	Up to 850	-[2]	2	2
Fluidyne Engineering Corp. 3900 Olson Memorial Hwy. Minneapolis, MN 55422	No	-	Wt	Fx	8-70	7-50	15-650	Up to 750	-[2]	1	1
Foster Wheeler Boiler Corp. 110 S. Orange Ave. Livingston, NJ 07039	Yes	Solids Circulation Systems, Inc.	Wt	Fx, Fcb	48-930	40-600	150-2400	Up to 1050	-[2]	8	11
International Boiler Works Co. 36 Birch St. E. Stroudsberg, PA 18301	No	-	Wt	Fx, Fcb	2.5-135	2-100	15-700	250-650	-[2]	4	4
Johnston Boiler Co. 300 Pine St. Ferrysburg, MI 49409	Yes	Combustion Systems, Ltd.	Wt, Ft	Pcb	30-100	25-70	15-860	Up to 750	-[2]	19	29
E. Keeler Co. 238 West St. Williamsport, PA 17701	No	-	Wt	Fx	48-290	40-200	100-800	Up to 800	-[2]	1	1

TABLE 3.2-6. DOMESTIC AFBC MANUFACTURERS[35] (Continued)

Company Address	AFBC Boiler Technology		Watertube or Firetube Boiler	Types of FBC Systems Offered	Boiler Capabilities Commercially Available					Number of Units Installed	
	Built Under License	Licensing Company			Heat Input,[1] $\times 10^6$ Btu/hr	Steam Capacity, x1000 lb/hr	Pressure psig	Temperature, °F	Fuel(s)	USA	Total
Pyropower Corp.[8] P. O. Box 81608 San Diego, CA 92041	No	-	Wt	Fcb	60-590	50-400	200-2500	Up to 950	-[2]	1	1
Riley Stoker Corp. 9 Neponset St. Worcester, MA 01606	Yes	Fluidized Combustion Contractors Ltd.	Wt	Fx	More than 48	More than 40	150-2600	Up to 1005	-[2]	0	0
Solids Circulation Systems, Inc. P. O. Box 2325 Boston, MA 02107	No	-	Wt	Fcb	24-285	20-200	150-1800	Up to 850	-[2]	0	0
Struthers Wells Corp. 1103 Pennsylvania Ave. W Warren, PA 16365	Yes	Battelle Memorial Institute	Wt	Fcb[9]	60-360	50-250	Up to 2650	Up to 900	Coal Petroleum coke Lignite	2	2
Sulzer Brothers, Inc. 200 Park Ave. New York, NY 10017	No	-	Wt	Fx	24-155	20-100	145-1450	350-977	Coal	0	1
Wormser Engineering, Inc. 225 Merrimac St. Woburn, MA 01888	No	-	Wt	Fx[10]	12-140	10-100	15-1000	Up to 750	-[1]	2	2
York-Shipley, Inc. P. O. Box 349 York, PA 17403	No	-	Ft	Fx	3.6-110	3-90	15-300	250-421	Coal Wood-waste Biomass	12	12

Footnotes:

1. Estimated assuming saturated feedwater at 10 psig and boiler efficiency of 82 percent
2. Designed to burn the following fuels separately or in combination: coal, wood-waste, biomass, liquid wastes or sludges, coal-washing wastes.
3. Combination firing has limitations depending on the type of fuel burned.
4. Designed to meet customer requirements.
5. In conjunction with Energy Resources Co.
6. In conjunction with E. Keeler Co. and Dorr-Oliver, Inc.
7. In conjunction with E. Keeler Co. and Curtiss-Wright Corp.
8. Pyropower is jointly owned by A. Alhstrom Oly (Finland) and General Atomic (U.S.).
9. Combuster included a dense bed section to enhance reactivity.
10. Multistage fluidized bed.

Abbreviations:

Fcb--Full circulating bed

Ft--Firetube boiler

Fx--Fixed (bubbling) bed

Pcb--Partial circulating bed

Wt--Watertube boiler

wide variety of applications. The guarantees offered vary by vendor, but can cover performance in areas such as steam quality and quantity, emissions, and combustion efficiency.

3.2.4 Summary of Existing and Planned AFBC Units

A summary of the existing and planned sites of domestic coal-fired AFBC units is listed in Table 3.2-7. Foreign coal-fired AFBC units, domestic and foreign alternate fuel and multifuel AFBC units are listed in Appendix E. The majority of the AFBC units are based on the conventional bubbling bed design, with a few units based on the circulating bed design. Only two units have staged beds. The sites listed range in size from 2,500 to 352,000 lb/hr of steam at pressures of up to 2650 psig (3 to 182 MMBtu/hr). Over twenty different types of fuel, including low-rank fuels (lignite and peat) and wastes from agricultural and municipal sectors and process industries, are burned. In addition to the units listed, there are over 2000 AFBC boilers in China.[35] These boilers are generally small and burn low grade fuels containing up to 70 percent ash.

Of the 80 AFBC sites in the United States, coal is the only design fuel in 14 units and is one of several design fuels in 9 units. It should be recognized that AFBC units constitute only a very small portion of the total domestic operating industrial boiler population.

Excluding AFBC boilers that were test, demonstration, undisclosed, or uncompleted units, eight AFBC boilers in the United States were identified which burn coal either alone or as one of several fuels. The operators of these AFBC boilers were contacted to obtain specific information concerning the operation of and emissions from these boilers. (The operating parameters of test and demonstration units, along with test results, are well documented in literature.) Seven responses were received. One operator indicated that their AFBC boiler was only a backup unit, and, although it was capable of firing coal, oil and natural gas were the primary fuels. Another operator has just brought an AFBC boiler on line after a series of serious equipment problems. Therefore, information concerning boiler performance was not available. The information collected from the

TABLE 3.2-7. EXISTING AND PLANNED DOMESTIC COAL-FIRED AFBC UNITS[35]

Plant Owner	Location	Heat Input[1] x10^6 Btu/hr	Steam Capacity, x1000 lb/hr	Steam Pressure psig	Steam Temperature °F	Design Fuel (s)	Manufacturer	Type of Project	Type of Financing	Commercial Service Date
Tennessee Valley Authority	Paducah, Kentucky	182	120[2]	2400	1000	C	BW	D	P/G	6/82
Georgetown University	Washington, D.C.	120	100	275	Sat	C	FWC	D	P/G	1/79
Iowa Beef Processors, Inc.	Amarillo, Texas	90	70	650	550	C	WOR[3]	Com	P	7/82
Idaho National Eng. Lab.	Idaho Falls, Idaho	82	68[4]	150[5]	Sat[5]	C	FWC	Com	P	12/83
Kentucky Agricultural Energy Corp.	Franklin, Kentucky	73	60[4]	550	Sat	C	FWC	Com	P	10/82
Central Ohio Psychiatric Hospital	Columbus, Ohio	72	60	150	Sat	C	FCL	Com	G	NAv
Gulf Oil Exploration & Prodn. Co.	Bakersfield, California	54	50	2500	Sat	C	PYR	Com	P	1/83
Texas Tar Sands Ltd.	Maverick City, Texas	54	50	2500	-	C	ERC[6]	Com	P	12/82
U.S. Navy	Great Lakes, Illinois	66	50	365	560	C	CEP	D	P/G	9/81
Van Buren County Alcohol, Inc.	Bonaparte, Iowa	24	20	225	Sat	C	DED	Com	P	8/81
Babcock & Wilcox Co.	Alliance, Ohio	32	20	150	1000	C	BW	D	P	5/78
Lowell Technological	Lowell, Massachusetts	24	20	125	325	C	WOR	Com	G	6/83
School Heating	Spencer, Indiana	2.9	2.5[4]	30	Sat	C	JBC	NAv	NAv	11/82
Manufacturing Plant	Fortville, Indiana	3.0	2.5	150	Sat	C	JBC	NAv	NAv	1/83

Footnotes:

1. Estimated assuming saturated feedwater at 10 psig and a boiler efficiency of 82 percent.
2. Initial rating; 190,000 lb/hr in the future.
3. In conjunction with International Boiler Works Co.
4. Two units installed.
5. Future steam conditions are 650 psig/750°F.
6. In conjunction with C-E Natco, a division of Combustion Engineering, Inc.

Abbreviations:

C--Coal
Com--Commercial project
D--Demonstration project
G--Government financing
NAv--Not available
P--Private financing
P/G--Private/government financing
Sat--Saturated

Manufacturers:

BW--Babcock & Wilcox Co.
CEP--C-E Power Systems, a division of Combustion Engineering, Inc.
DED--Dedert Corp., Thermal Process Division
ERC--Energy Resources Co.
FCL--Fluidized Combustion Contractors Ltd.
FWC--Foster Wheeler Boiler Corporation
JBC--Johnston Boiler Co.
PYR--Pyropower Corp.
WOR--Wormser Engineering Co.

five remaining operators is summarized in Table 3.2-8. Comparison of the units indicates the variability in the design and operating conditions for these initial commercial installations.

Plant A utilizes a circulating bed design with staged combustion air. In addition to the solids recycle provided by the circulating bed, the capability exists for recycle of solid materials collected from the flue gas downstream of the circulating bed. However, the operator does not believe that the benefits derived from this additional solids recycle are worth the trouble associated with its use. One benefit that has been previously identified, which solids recycle provides, is the reduction in the amount of limestone required to reduce SO_2 emissions to a specific level. Since this plant is located near a limestone quarry, the Ca/S ratio ($\sim$3.5) is varied as needed to achieve 90 percent SO_2 removal without significant concern for limestone usage. The fuel consists of varying combinations of coal containing 0.5 percent sulfur and petroleum coke containing about 7 percent sulfur. An average fuel combination contains approximately 2 percent sulfur. Compliance testing has been completed, but the data are not yet available.

An AFBC boiler with a circulating bed design has been constructed at Plant B. It is equipped with staged combustion air. Operation began in mid-July, but data are not yet available. The unit is currently burning coal with a 0.6 percent sulfur content. Possible future fuels include petroleum coke and oil-impregnated diatomaceous earth.

Plant C features a conventional bubbling bed with solids recycle. A Ca/S ratio of 2.0 is currently being used during the shakedown phase, but the SO_2 removal for this ratio has not yet been determined. As of the date of contact, the longest continuous operating period was four hours. After continuous operation is attained, the operating conditions will be adjusted to satisfy environmental regulations. One of two available coals, containing 0.8 percent and 1.5 percent sulfur, will be burned depending on cost considerations.

Plant D has a conventional bubbling bed without solids recycle or staged combustion air. Limestone is used only as a bed material (i.e., not

TABLE 3.2-8. SUMMARY OF INDUSTRIAL COAL-FIRED AFBC BOILER OPERATOR CONTACTS*

	Plant A	Plant B	Plant C	Plant D[1]	Plant E
Construction	Field	Field	Field	Field	Package
Bed Configuration	Circulating	Circulating	Conventional Bubbling Bed	Conventional Bubbling Bed	Conventional Bubbling Bed
Heat Input,[2] 10^6 Btu/hr	54	54	54	24	48
Features					
Solids Recycle	Yes[3]	Yes	Yes	No	Yes[4]
Staged Combustion Air	Yes	Yes	No	No	No
Limestone for SO2 Removal	Yes	Yes	Yes	No[5]	No
Recycle Ratio	NA[6]	Not Determined	Not Determined	NA	NA
Primary/Stoichiometric Air Ratio	0.6	Confidential	NA	NA	NA
Ca/S Ratio	3.5	3 or 4	2	NA	NA
Percent SO_2 Removal	90	Not Determined	Not Determined	NA	NA
Fuel					
Type	Coal	Coal	Coal	Coal	Coal
Heating Value (HHV), Btu/lb	7,937	10,000	Not Available	Not Available	12,085
Sulfur Content, Percent	0.5[7]	0.6	0.8/1.5[8]	1.0	3
Alternate Fuels	Petroleum Coke	Coke[9]	–[10]	None	None
Boiler Efficiency, Percent	72	Not Determined	Not Determined	Not Available	83.5
Availability, Percent	85[2]	Not Determined	Not Determined	Not Available	
CEM Equipment					
SO_2	Yes	Yes	Yes	No	No
NO_x	Yes	Yes	Yes	No	No
CO	Yes	No	Yes	No	No
CO_2	Yes	No	No	No	No
Particulates	Yes	Yes	Yes	No	No
Recurring Problems	None	NA	NA	NA	Water Tube and Wall Erosion
Status	Operational Dec. 1981. Compliance testing completed July 1983.	Operational July 1983.	Operational Aug. 1983.	Operational Aug. 1981. Currently operating with cost-cutting measures.	Operational Apr. 1980. Problems with erosion of water tubes and walls.

* Footnotes located on next page.

FOOTNOTES FOR TABLE 3.2-8.

[1] Information gathered from manufacturer at suggestion of operator.

[2] Estimated assuming saturated feedwater at 10 psig and a boiler efficiency of 82 percent.

[3] Additional solids recycle, beyond that provided by the circulating bed, is available but not being used.

[4] Solids recycle incorporated in original unit but presently inoperable due to mechanical problems.

[5] Limestone used only for bed material due to liberal emission requirements and as a cost-cutting measure.

[6] Not applicable.

[7] Total fuel stream. Petroleum coke (alternate fuel) contains ~7 percent sulfur and coal contains ~0.5 percent sulfur.

[8] Two coals with different sulfur contents will be used.

[9] After unit has begun operation, oil-impregnated diatomaceous earth will be tested for use as a fuel.

[10] The decision to use or not use alternate fuels has not been made.

in sufficient quantities to remove a significant amount of SO_2) due to less stringent SO_2 emission requirements and as a cost-cutting measure. Information concerning SO_2 emissions and environmental regulations was not available. The plant has burned a variety of coals. A 1.0 percent sulfur coal is the current fuel.

Plant E features a conventional bubbling bed. Solids recycle was originally available but is not currently operable due to mechanical problems. No effort is being made to control SO_2 emissions. Major boiler modifications and additional material handling systems would have to be installed before limestone could be used to control SO_2 emissions. The current fuel is a 3 percent sulfur coal.

3.2.5 Recent Improvements and Technology Trends

Several modifications to and deviations from the traditional bubbling bed AFBC technology have been reviewed, including solids recycle, staged combustion air, staged beds, and circulating bed configurations. Research in these areas has resulted in improved system designs and has defined the direction of FBC technology development. Also, ongoing and near-term research involving the environmental characterization of advanced FBC designs is expected to result in more optimized performance. These issues are reviewed in the following discussion.

3.2.5.1 Design Configurations -- At the present time, none of the various design configurations dominates the emerging AFBC industrial boiler market. Boilers featuring the various designs have recently been installed in a variety of applications, although continued commercialization may favor certain designs over others or specific designs for certain applications (e.g., circulating bed technology for enhanced oil recovery steam generation). Because the various design configurations have yet to be completely optimized, it is not presently known which design(s) will emerge as the next generation of widely accepted commercial technology. Therefore, research directed towards environmental characterization of future AFBC technology must, at this time, focus on all of the various commercial design configurations. Significant commercial installations representing these designs include:

- Gulf Oil Exploration and Production Company, Bakersfield, California -- a Pyropower circulating bed design scheduled for startup in late 1983 for steam generation in an enhanced oil recovery (EOR) application;

- Conoco, Inc., Uvalde, Texas -- a Battelle/Struthers Wells Corporation circulating bed design started up in early 1982 for steam generation in an EOR application (the unit, which is designed to fire coal or a mixture of coal and petroleum coke, is scheduled for optimization studies);

- Iowa Beef Processors, Inc., Amarillo, Texas -- a Wormser staged bed design started up in late 1982;

- Lowell Technological Institute, Lowell, Massachusetts -- a Wormser staged bed design scheduled for startup in mid-1983;

- Texas Tar Sands, Limited, Maverick County, Texas -- a more traditional AFBC design featuring solids recycle (Energy Resources Company) started up in late 1983; and

- Kentucky Agricultural Energy Corporation, Franklin, Kentucky -- a traditional AFBC design with solids recycle (Foster Wheeler Boiler Corporation) started up in late 1982.

These installations represent, from a technical standpoint, state-of-the-art candidates for environmental characterization studies.

In addition to investigating existing AFBC designs, government support of higher-risk innovative FBC concepts, such as PFBC for industrial applications, staged cascade designs, and ultra-high velocity combustion units, is expected to continue.[21]

3.2.5.2 Environmental Characterization -- A key advantage of FBC technology over conventional coal combustion technology is the ability of FBC to provide in-situ control of SO_2 and NO_x emissions. Ongoing and future research and development efforts are and will be focused on further defining the interrelationships between emissions control and boiler performance.

TVA has targeted an SO_2 control level of 90 percent at a Ca/S ratio of 2.0 for FBC units in utility applications.[36] In addition to the new design configurations previously reviewed, substantial progress towards approaching this target performance level has resulted from extensive investigation of SO_2 retention mechanisms as well as research designed to optimize sorbent selection and utilization.[34,37] Additional concepts designed to improve SO_2 control or sorbent utilization, such as salt addition and sorbent regeneration, have received and are expected to receive considerable emphasis from various investigators. However, these concepts are unlikely to gain acceptance among potential industrial users in the near-term due to the costs and/or risks involved.

TVA has targeted a performance level for NO_x emissions from utility FBC units of 0.2 $lb/10^6$ Btu.[36] Recent research has emphasized NO_x control in conjunction with SO_2 control and combustion efficiency improvement. In the past, testing has tended to focus more on optimizing SO_2 control and combustion efficiency than on minimizing NO_x emissions. Also, fundamental investigation of NO_x formation and reduction mechanisms is expected to result in a better understanding of the relationships between NO_x emissions, SO_2 emissions, and combustion efficiency.

Historically, control of particulates from AFBC boilers has been accomplished through the use of conventional technologies -- cyclone collection followed by fabric filtration or electrostatic precipitation. However, fly ash from FBC boilers has been recognized to be markedly different in composition from that emitted from conventional boilers. In particular, FBC ash contains greater amounts of carbon and calcium and lesser amounts of sulfur-bearing compounds. This non-conventional composition poses resistivity problems for ESPs and fire hazards for fabric filters.[35] Nonetheless, the use of conventional particulate control technologies for industrial FBC boilers is expected to continue, and optimization of their performance is expected to occur as the degree of research and demonstration accelerates.

The solid waste material from FBC units has received considerable research attention in the past, particularly with regard to its potential use as a marketable by-product (e.g., as structural material or as an agricultural supplement). Ongoing and future research efforts may pursue this topic, but it is expected that a significant amount of work will also be aimed at the environmental impacts associated with disposal of the waste by more traditional methods.[38,39,40] One important issue is the Resource Conservation Recovery Act (RCRA) classification of AFBC solid waste. Toxicity characteristics are a potential concern, but recent investigations have shown that FBC waste would typically be classified as nonhazardous, according to RCRA provisions.[41,42] However, laboratory studies have indicated high levels of pH, total dissolved solids (TDS) content, and sulfate content in leachate from FBC waste.[43]

These issues are discussed further in Section 4.

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SECTION 4
SYSTEM PERFORMANCE DATA

The effects of specific AFBC operating conditions and design configurations on SO_2, NO_x, and PM emissions are examined in this section. Recent data correlating emissions control to process design and operating variables are presented. Most of these data were obtained from test-scale AFBC units. Data available from commercial operating facilities, although limited, are also presented. Finally, other factors affecting boiler performance are reviewed.

4.1 SUMMARY OF SO_2 EMISSION DATA

The ITAR discussed the relationship between the level of SO_2 emissions from an AFBC boiler and the following design and operating variables:[1]

- Sorbent particle size,
- Sorbent reactivity,
- Gas residence time,
- Bed temperature,
- Feed mechanisms,
- Excess air, and
- Ca/S molar feed ratio.

The effect of these variables will be briefly reviewed.

Modifications to the conventional bubbling bed have resulted in the following design concepts which also affect SO_2 emissions:

- Solids recycle,
- Staged combustion,
- Staged beds, and
- Circulating beds.

Performance data will be presented which demonstrate the effect of these designs on SO_2 emissions.

Also presented in this subsection is information related to (1) the effect of coal characteristics on SO_2 emissions, (2) enhanced sulfur capture methods, and (3) SO_2 emissions control data for the different design configurations.

4.1.1 Design and Operating Variables Affecting SO_2 Emissions

Limestone utilization increases as the particle size decreases. Tests on various limestone grain sizes have shown that sulfur capture drops off rapidly, from 85 percent to 65 percent, as grain size increases from 400 μm to 1000 μm.[2] The increased sulfur capture is attributed to the increased surface area per unit mass of limestone.

Limestone reactivity is also affected by the calcined limestone's pore size and chemical constituents besides calcium.[3] Calcined limestone with large pores tends to be more fully utilized. Small pores have more surface area per unit mass and allow for faster initial reaction between SO_2 and sorbent, but they tend to plug quickly with sulfate. The presence of $MgCO_3$ causes a slightly different grain structure which provides greater pore surface area resulting in higher limestone utilization. Sodium present in limestone has also been shown to increase limestone utilization.[3]

Another variable which affects SO_2 emissions is gas residence time. Gas residence time is the time period required for a unit volume of gas to pass through the bed and is defined as the ratio of the expanded bed height to the superficial velocity. As gas residence time increases, SO_2 removal efficiency improves due to the increased time available for calcination and sulfation reactions.[4] The ITAR identified a critical gas residence time, 0.6 to 0.7 seconds, below which SO_2 removal was significantly reduced.

The bed temperature directly affects the efficiency of sulfur removal. A temperature of at least 1400°F is necessary to fully calcine the limestone and form CaO, the reactive form of the sorbent. Early research referred to in the ITAR indicated an optimum bed temperature for SO_2 removal of between 1400° and 1600°F, depending on the coal and sorbent in use and on the specific operating parameters. More recent research supports this temperature range.[5,6,7]

Removal of SO_2 can be affected by the coal and limestone feed points and feed system. Overbed feed systems tend to be simpler and more reliable. However, SO_2 released above the bed, where sorbent is not available for SO_2 capture, is a potential problem. In addition, limestone fines fed above the bed may be elutriated from the system before being utilized. Recycle of elutriated material is recommended when overbed feeding is employed. Testing of overbed feeding is planned at the TVA/EPRI 20 MWe pilot plant.

Underbed feed mechanisms provide longer bed residence times for coal and limestone particles, increasing combustion efficiency and limestone utilization. However, underbed feed designs tend to be more complex and expensive and less reliable. Earlier underbed feed systems design guidelines relied on a feed point every 9 square feet.[8] The TVA/EPRI 20 MWe pilot plant was designed to require fewer feed points (1 feed point/18 square feet). One of the major problems encountered to date at the TVA/EPRI pilot plant has been erosion of the underbed feed lines.[9]

The excess oxygen level also has an effect on SO_2 removal, as stated in the ITAR. Recent tests have confirmed that an increase in excess air increases SO_2 removal. In one research program, SO_2 removal increased from 87.5 to 96 percent as the air ratio (combustion air to stoichiometric air) was increased from 0.8 to 1.25.[10]

The Ca/S molar feed ratio has the greatest impact on SO_2 emissions. Figure 4.1-1 represents recent test data from five conventional AFBC units without solids recycle.[6,9,11,12,13] It can be seen that as the Ca/S ratio

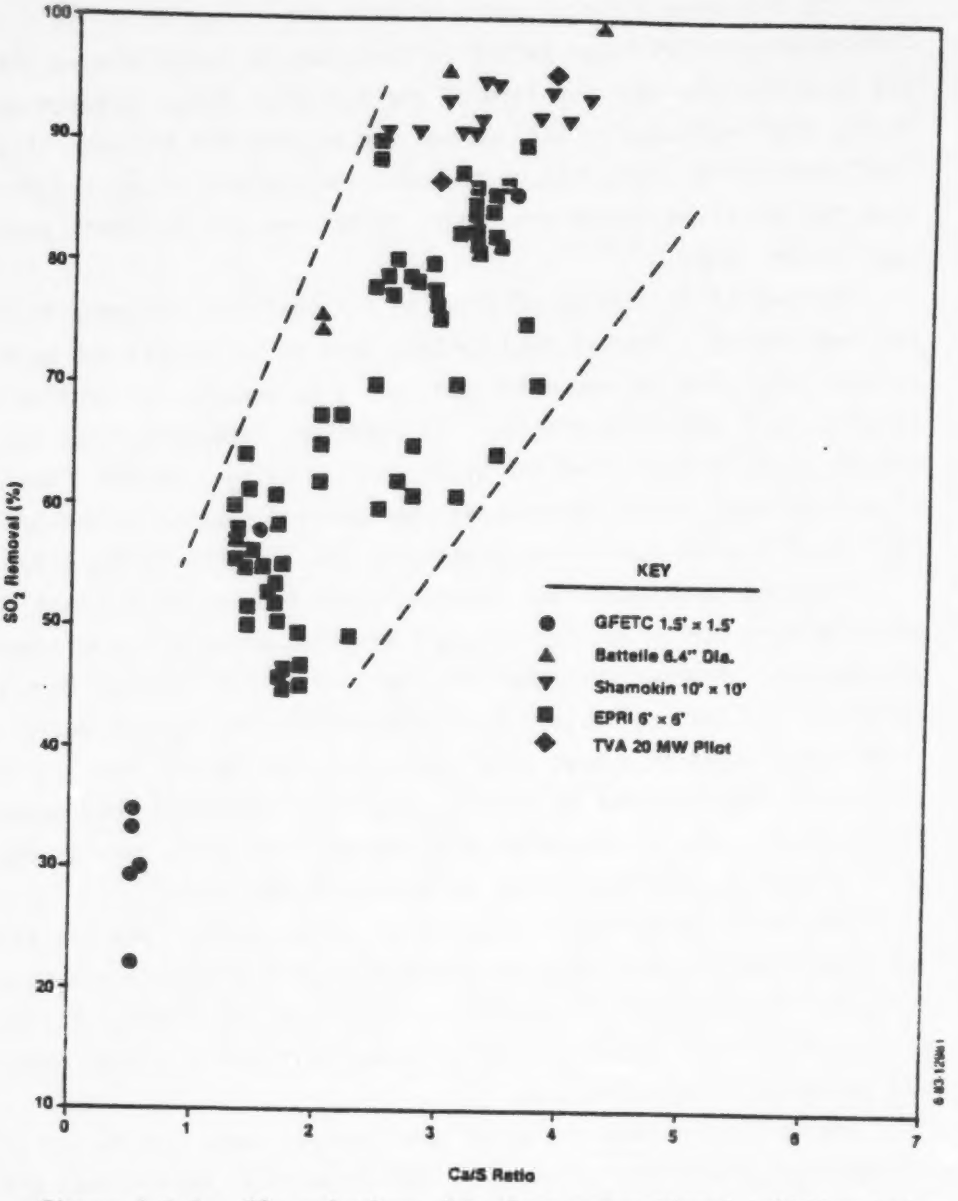

Figure 4.1-1. SO_2 emissions data from conventional bubbling bed AFBC units without solids recycle.(6,9,11,12,13)

increases, SO_2 removal increases. The data are somewhat scattered due to the effects of other variables that affect SO_2 emissions. However, the general trend is still apparent. While the majority of the data are from the EPRI 6'x6' unit, the results from other units show the same trend when their data are examined independently. These data show no significant deviation from earlier experimental data presented in the ITAR.

4.1.2 Solids Recycle

The recycle of elutriated bed material can have a significant effect on SO_2 removal at a set Ca/S ratio since the recycled material typically contains unreacted sorbent. Figure 4.1-2 is a summary of SO_2 removal data for several different conventional bubbling bed AFBC units which incorporate recycle of elutriated material.[9,13,14,15] When compared to the SO_2 emissions data from traditional units without recycle (presented earlier in Figure 4.1-1), the general trend for solids recycle to lower the required Ca/S ratio to achieve a specific level of SO_2 removal is apparent. The scatter in the data results from the different operating conditions of the various units represented in the figure. All of the data from the METC 18" unit and some of the data from the Johnston Test Unit were collected at bed temperatures from 1425° to 1500°F. These data indicate higher SO_2 retention levels than data from the EPRI and TVA units which operate at bed temperatures of approximately 1550°F. The remaining data from the Johnston Test Unit represent operation at higher temperatures and show decreased SO_2 removal. As stated in the ITAR, higher bed temperatures lower SO_2 removal. Therefore, bed temperature appears to be one identifiable operating condition which is responsible for the difference in the data. Other operating parameters such as sorbent reactivity, feed mechanism, and excess air could also be responsible for the variation in the data.

Tests to determine the effect of various levels of solids recycle on SO_2 emissions have been performed on two units. Figures 4.1-3 and 4.1-4 summarize recycle tests performed on the General Atomic 16" unit and the EPRI/B&W 6'x6' unit, respectively.[6,16] Data from both units demonstrate the beneficial effect of solids recycle on SO_2 retention. Recycle rate is

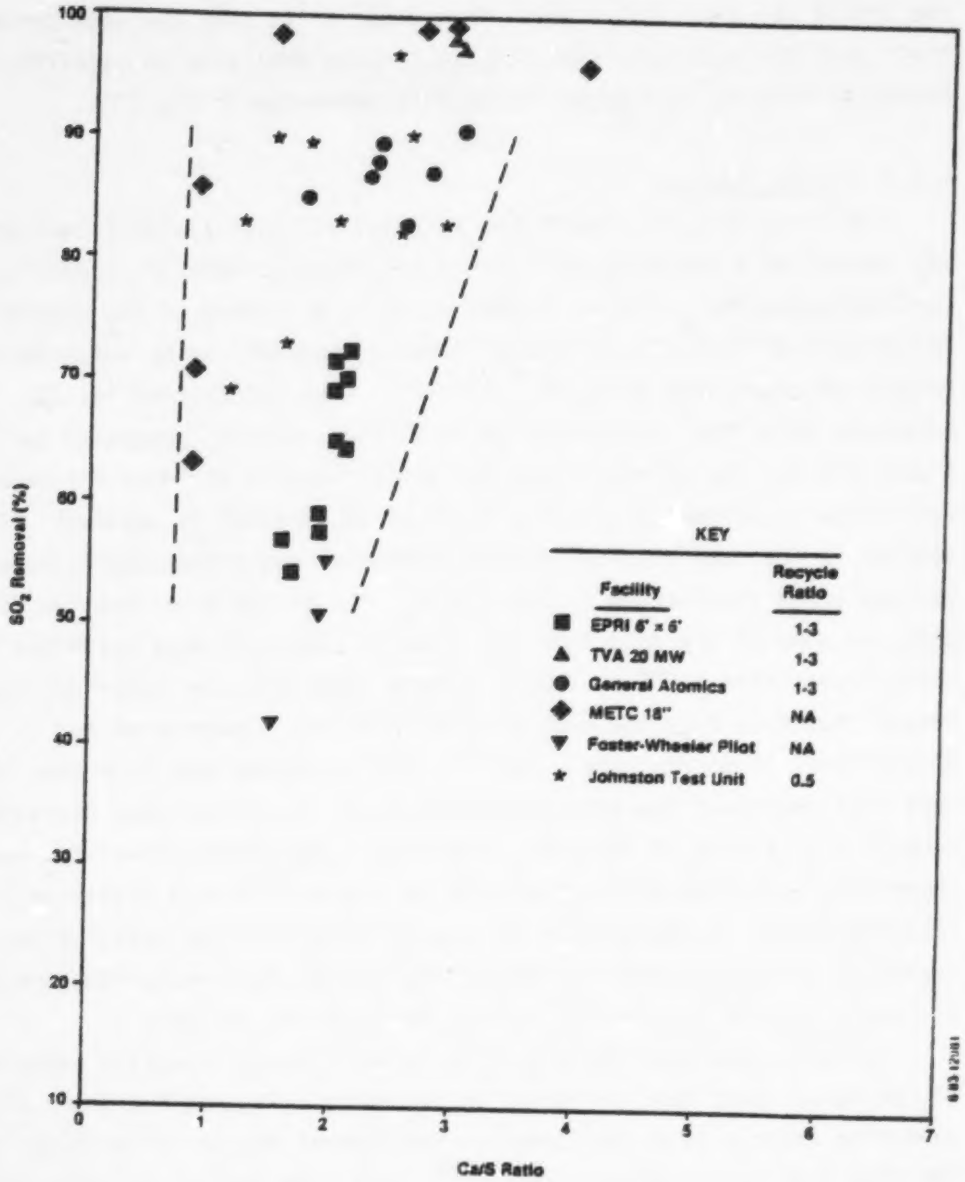

Figure 4.1-2. SO_2 emission data from conventional bubbling bed AFBC units with solids recycle.(9,13,14,15)

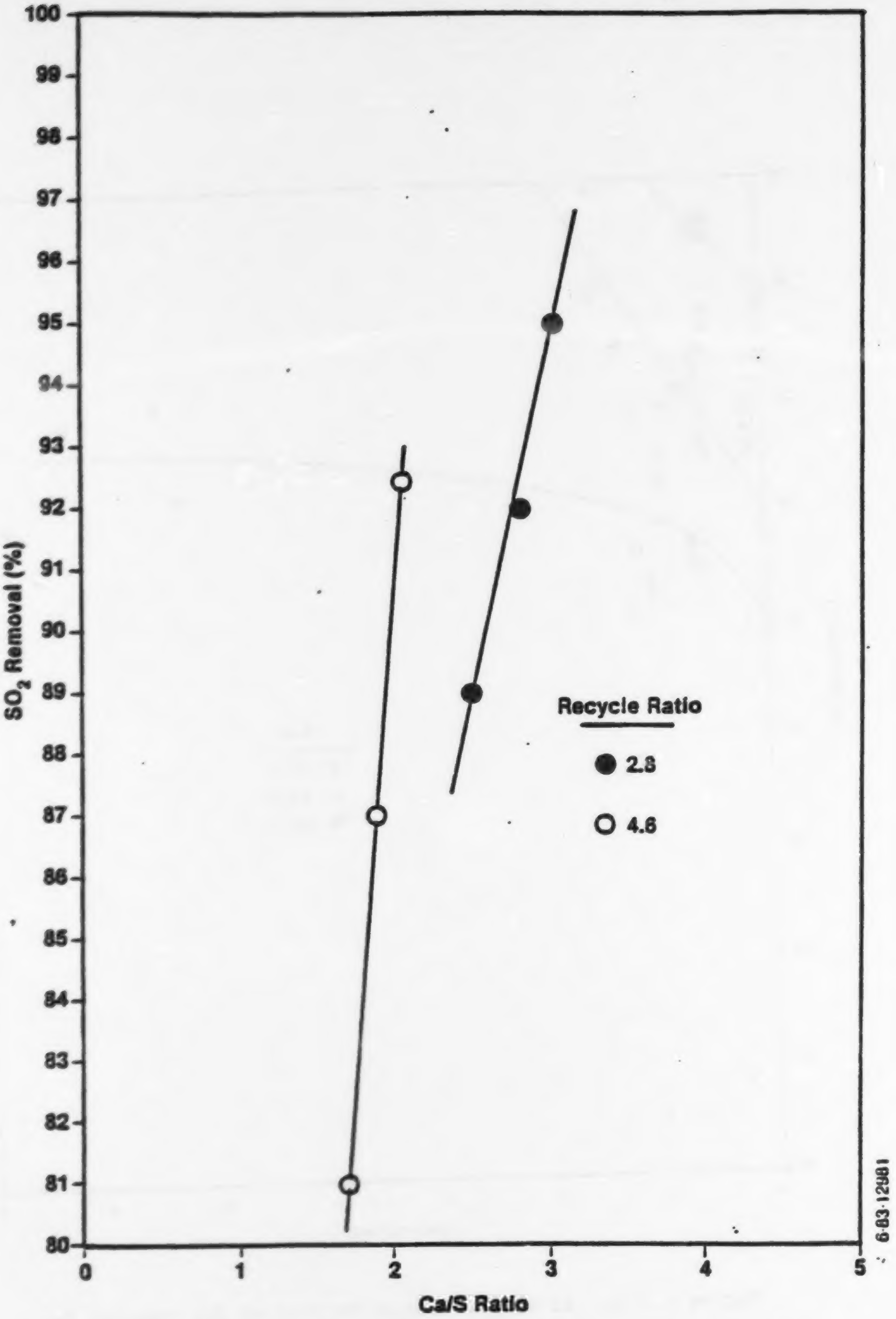

Figure 4.1-3. Effect of solids recycle on SO_2 removal for the General Atomic 16" test unit (16)

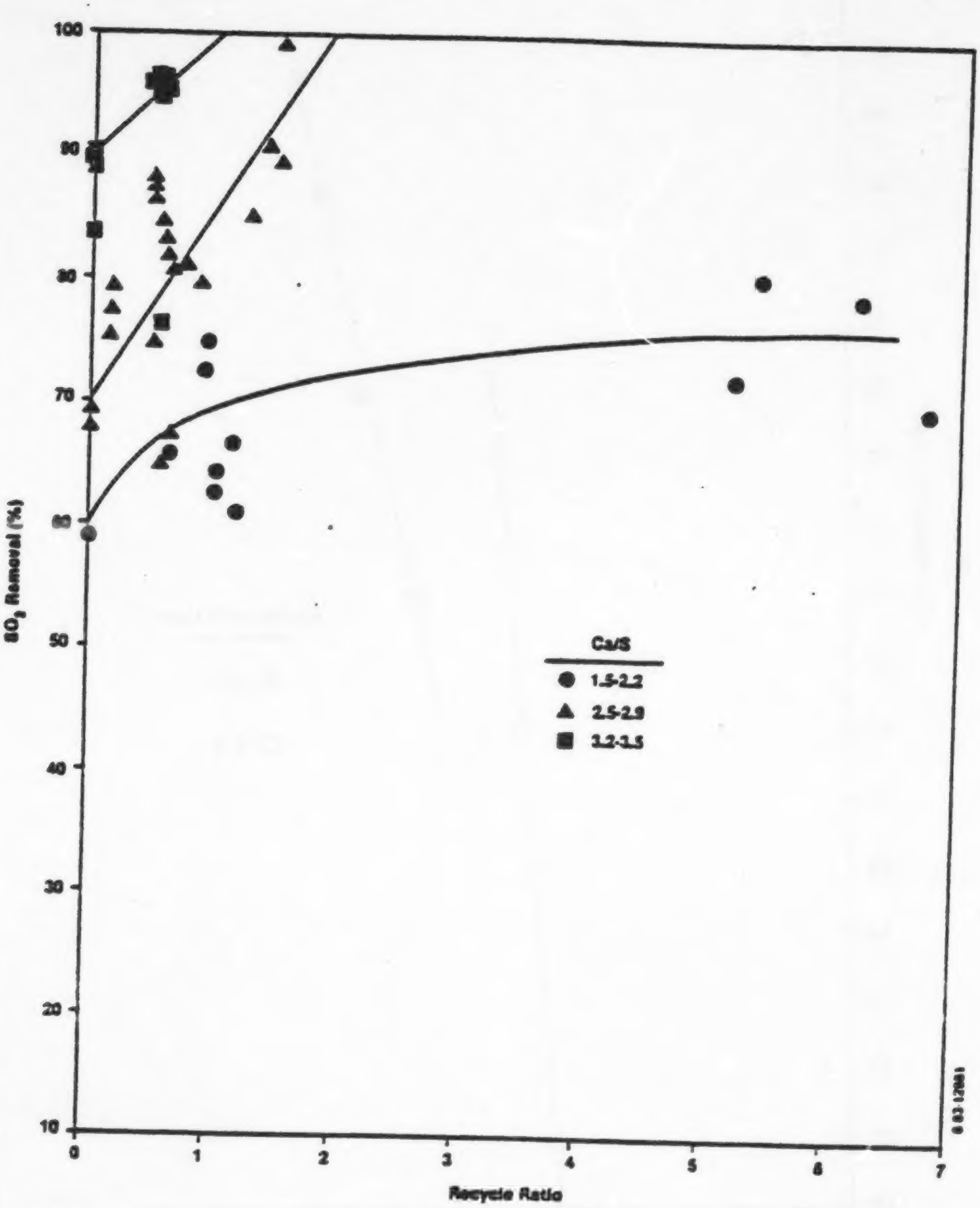

Figure 4.1-4. Effect of solids recycle on SO_2 removal for the EPRI/B&W 6'x 6' unit (6). (Curve fits obtained from literature source.)

defined as the mass flow rate of recycle solids divided by the coal mass feed rate. Higher recycle ratios result in improved SO_2 removal. However, the data from the EPRI/B&W unit indicated that for low Ca/S ratios (1.5-2.2) only moderate improvement results as the recycle ratio increases from 1.0 to 6.0. Testing conducted by the Argonne National Laboratory on several samples of recycled material from the EPRI/B&W 6'x6' unit provided some explanation for this phenomenon.[17] The study found that the ability of the recycled material to remove SO_2 was found to decrease quite rapidly as its degree of sulfation, defined as the ratio of sulfated calcium to total calcium, reached a 30 percent level.

4.1.3 Staged Combustion Air

Staging the combustion air is the primary method used to reduce NO_x emissions. (Refer to Subsection 4.2.3.) However, staging the combustion air creates a reducing zone in the bed which limits the extent of the $CaO-SO_2-O_2$ reaction that forms $CaSO_4$, resulting in slightly higher SO_2 emissions.[12] Figure 4.1-5 shows the effect of staged combustion air on SO_2 removal in the Battelle 6" test unit.[12] Recycle of elutriated material was not used for these tests.

The most important variable associated with staged combustion air is the primary air ratio, defined as the ratio of air introduced at the distributor plate to the stoichiometric air. The primary air ratio has an effect on SO_2 emissions. As the primary air ratio is lowered, SO_2 emissions are increased. Figure 4.1-6 demonstrates the effect of the primary air ratio on SO_2 removal for staged combustion air.[10] Sulfur removal is observed to drop off as the primary air ratio decreases to less than 1.0. Refer to Section 4.3 for discussion of SO_2/NO_x tradeoff.

4.1.4 Staged Beds

Combustion and desulfurization occur in separate beds in staged bed AFBC units. The SO_2 emission test results for three different limestones in the two-bed United Shoe Manufacturing Corporation (USMC) AFBC boiler are presented in Figure 4.1-7.[18] Although the staged bed design theoretically

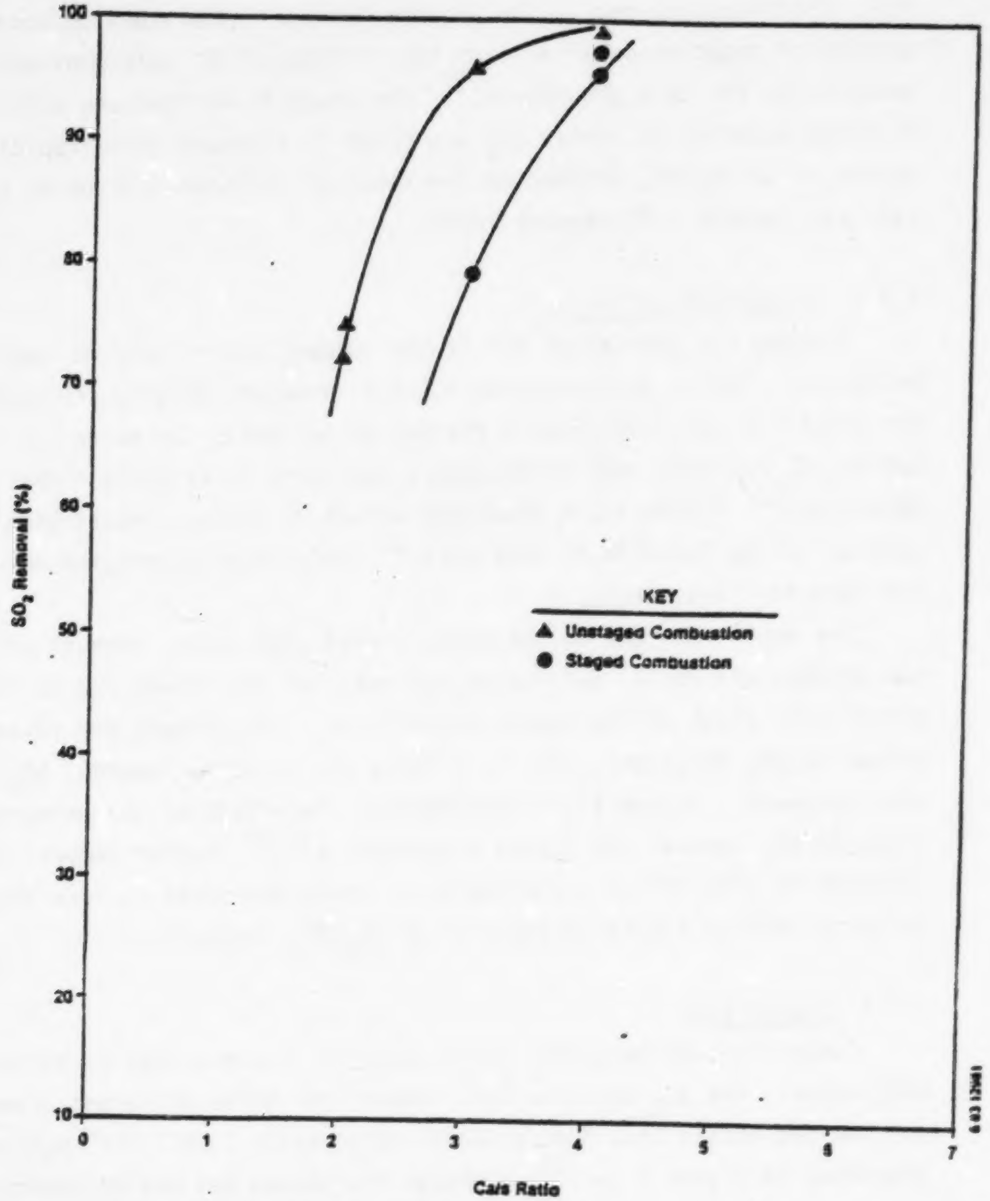

Figure 4.1-5. Effect of staged combustion air on SO_2 removal for the Battelle 6" test unit (12)

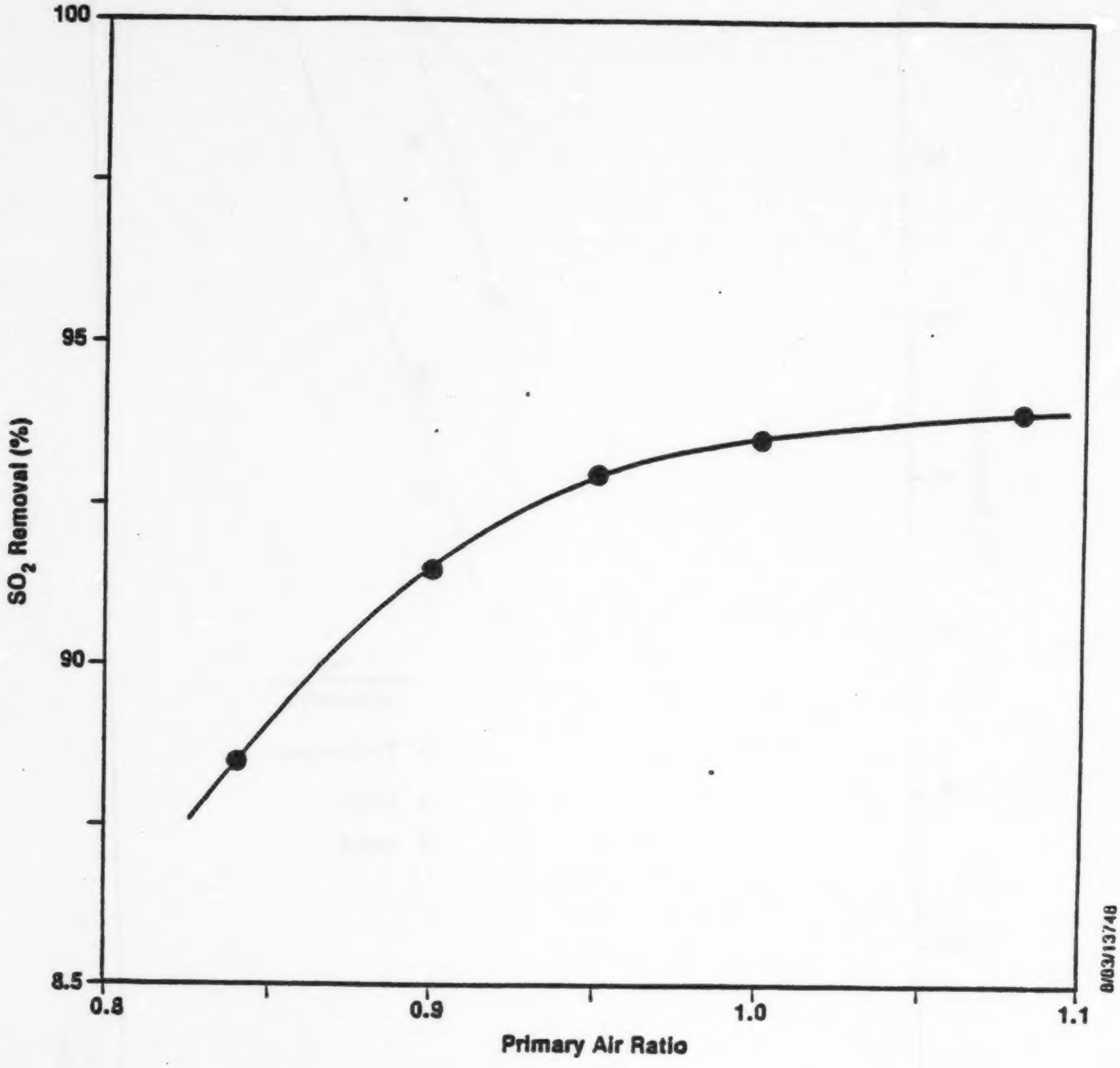

Figure 4.1-6. Effect of primary air ratio on SO$_2$ removal (10)

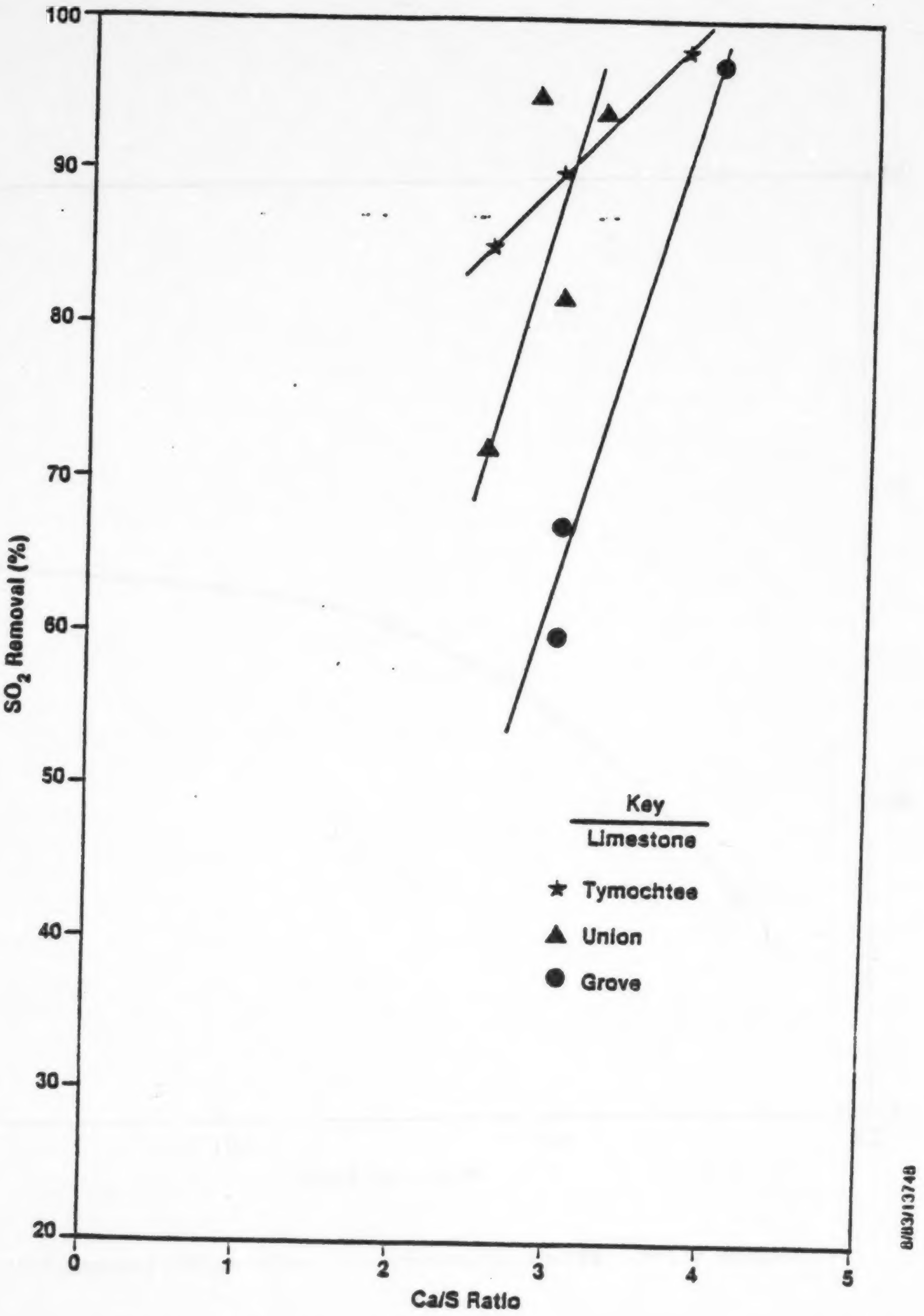

Figure 4.1-7. Staged bed SO_2 emission results for the United Shoe Manufacturing Corporation AFBC boiler (18)

provides an advantage for SO_2 removal, these data show no significant improvement in SO_2 removal efficiency for the staged bed design over the performance of conventional AFBC boilers.

4.1.5 Circulating Bed

Circulating bed AFBC units, which feature a recirculating entrained bed, have been demonstrated to achieve SO_2 removals of 90 percent with Ca/S ratios of 1.5.[19] Sulfur dioxide emission data for a Lurgi circulating AFBC boiler are presented in Figure 4.1-8.[19] These data support the superior SO_2 control levels achievable by a circulating bed design due to the solids recycle provided by the circulating bed. In the Battelle multi-solid AFBC, the entrained bed can be recycled to the combustion zone at different rates. The effect of the entrained bed recycle rate on SO_2 removal for a Battelle test unit is presented in Figure 4.1-9.[20] Sulfur removal is shown to increase with higher entrained bed recycle rates.

4.1.6 Coal Characteristics

Recent test data have indicated that coal characteristics can affect SO_2 emissions levels. In addition to sulfur content, factors such as the form of the sulfur and the alkalinity and quantity of ash affect SO_2 emissions. In addition, system reliability can be affected by the agglomerating tendencies of some coals containing high levels of sodium (e.g., lignites).

Tests conducted by Grand Forks Energy Technology Center and Morgantown Energy Technology Center on low-rank coals indicate that some lignites and low-sulfur subbituminous coals contain significant quantities of reactive calcium and sodium alkalinity in their ash. The tests were conducted on high-sodium and low-sodium lignites from a Beulah, North Dakota mine and on lignite from a San Miguel, Texas mine. The inherent alkali (calcium and sodium)-to-sulfur ratios were 1.20, 0.54, and 0.75 for the Beulah high-sodium, Beulah low-sodium, and the San Miguel lignites, respectively. To achieve 90 percent sulfur removal, the San Miguel lignite required additional limestone corresponding to an alkali-to-sulfur ratio of about

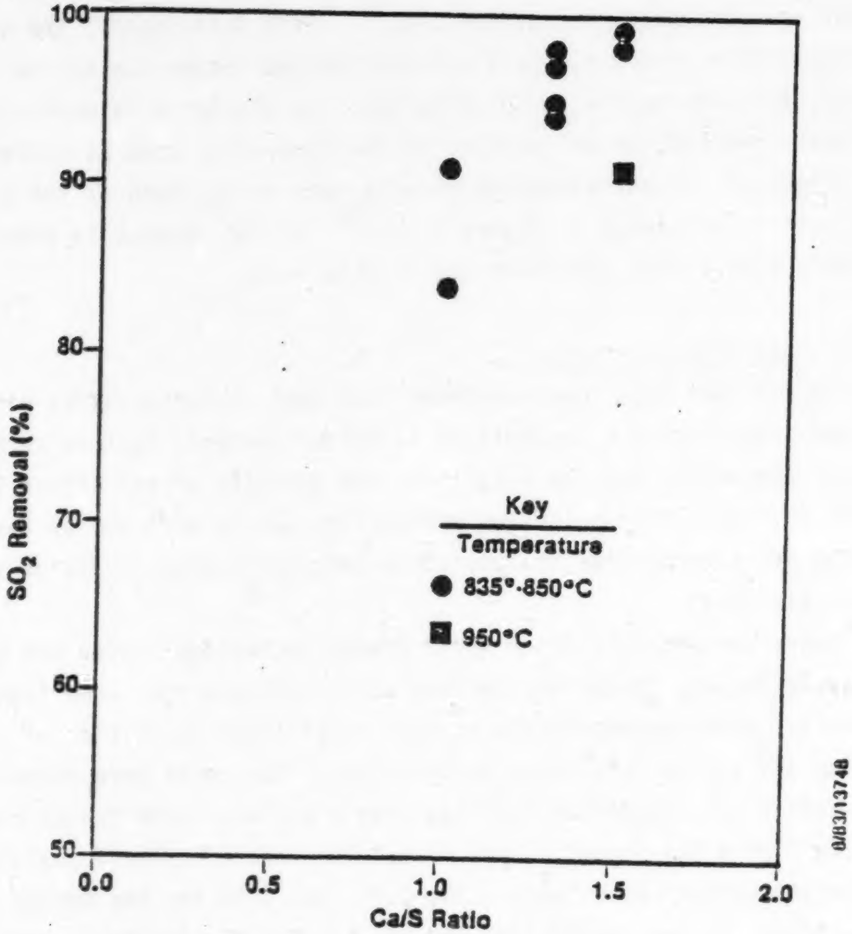

Figure 4.1-8. SO_2 removal test results for Lurgi circulating bed AFBC boiler (19)

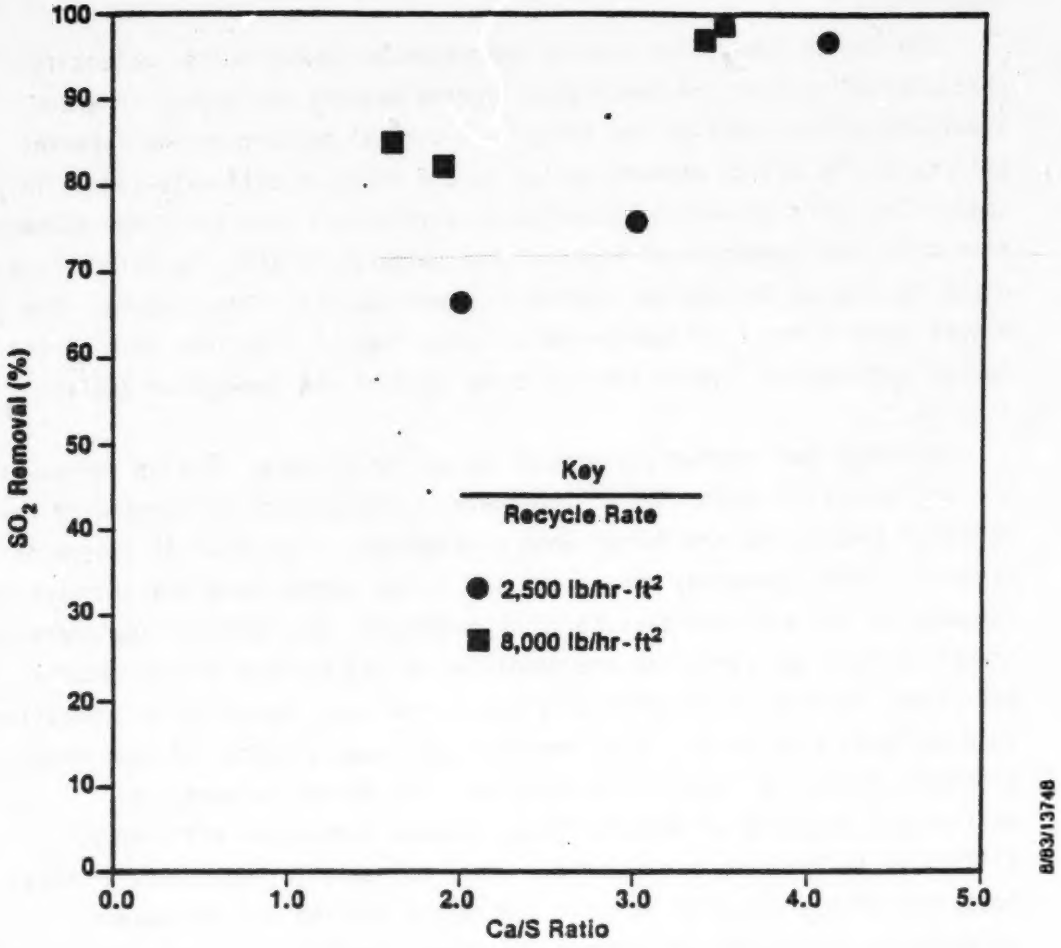

Figure 4.1-9. Effect of entrained bed recycle rate on SO_2 removal for a Battelle circulating bed AFBC test unit (20)

2.5. The Beulah low-sodium lignite required an added alkali-to-sulfur ratio of about 0.75 while the Beulah high-sodium lignite required no additional alkali to achieve 90 percent SO_2 removal. Figure 4.1-10 further illustrates the difference in the availability of the alkali to retain sulfur in the three coals.[5]

The Beulah low-sodium lignite demonstrated better sulfur retention characteristics than the San Miguel lignite despite the higher inherent alkali-to-sulfur ratio of the latter. A partial measure of the inherent ability of the ash to capture sulfur is the ratio of silica-to-sodium in the coal. The ratio of available sodium to available silica and other elements determines the formation of high-melting temperature alkali aluminosilicates which may tie up the sodium, making it unavailable for SO_2 capture. The San Miguel lignite has a silica-to-sodium ratio that is 4.8 times that of the Beulah high-sodium lignite and 1.3 times that of the low-sodium Beulah lignite.[5]

Although coal sodium contributes to sulfur capture, it also increases the agglomerating tendencies of the coals. Compounds or mixtures with low melting temperatures are formed when a relatively high level of sodium is present. These compounds reduce the ash fusion temperature and increase the tendency of the ash particles to stick together. Bed material agglomeration occurs as fuel ash particles are deposited on the surface of bed material particles, forming large solid clusters in the bed. Deposits on combustion zone surfaces also occur. Agglomeration can cause a number of operating problems, including loss of fluidization, loss of bed temperature uniformity, plugging of recycle lines, reduced combustion efficiency, difficulty in draining bed material, and a decrease in heat transfer rate. Emissions of SO_2 can also increase due to the coating and subsequent decrease in utilization of sorbent particles. Methods available to minimize agglomeration include bed flushing, operation at lower temperatures, operation with higher gas velocities, operation without recycle, and the addition of alkali suppressants.

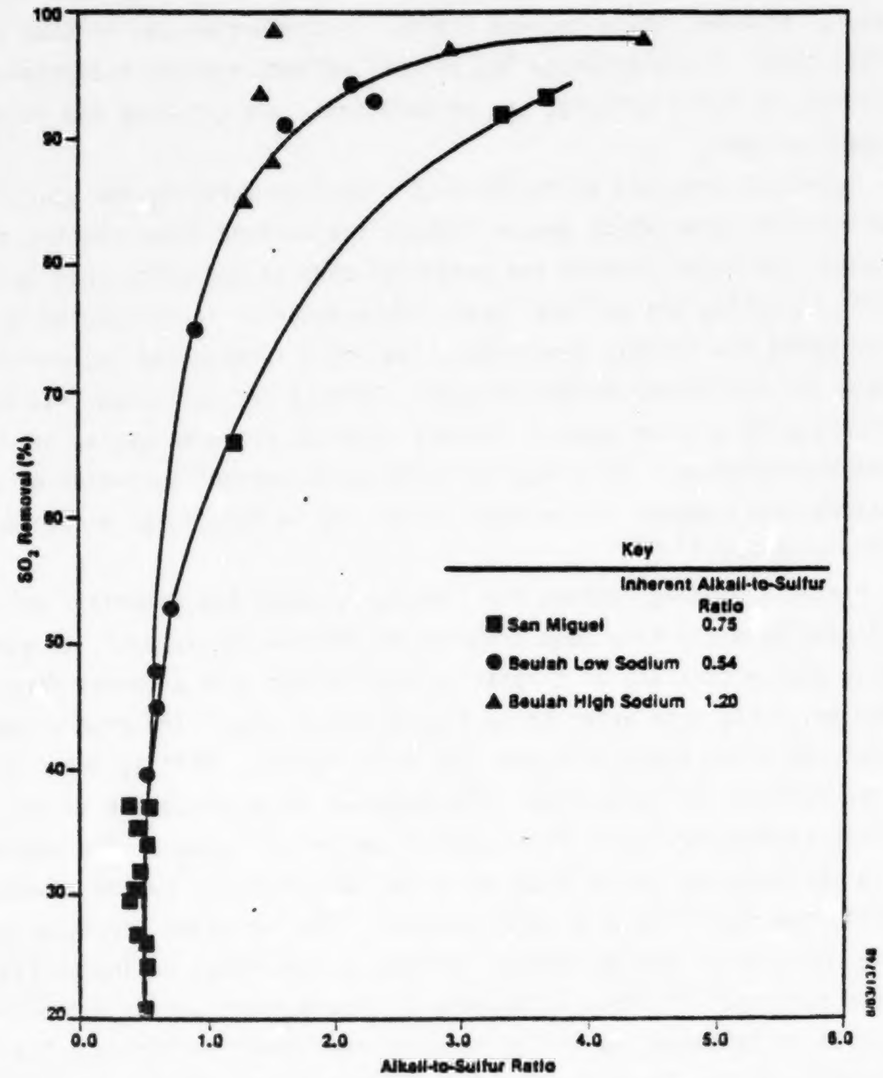

Figure 4.1-10. Effect of fuel content on SO_2 removal (5)
(Curves presented in literature source.)

4.1.7 Enhanced Sulfur Capture Methods

Recently, other methods have been investigated to provide enhanced SO_2 removal. However, these methods are not in commercial use in AFBC boilers at this time. These enhanced SO_2 removal methods include hydration enhanced sulfation, particle bonding, use of additives, and grinding and reinjection of spent sorbent.

Hydration enhanced sulfation (HES) involves spraying the spent bed material with water which passes through the sulfate layer coating the spent sorbent. The water reaches the unreacted core of CaO which then hydrates to $Ca(OH)_2$, cracking the sulfate layer. This material is reinjected to the boiler where the $Ca(OH)_2$ dehydrates, leaving a large-pored CaO particle exposed for additional sulfur capture. Testing has indicated that the optimum use of a given mass of sorbent consists of three cycles of sulfation/hydration. This results in 80 to 90 percent sulfation of calcium. Limestone requirements for an AFBC boiler may be reduced by a factor of two or more with HES.[21,22]

Particle bonding methods for limestone, spent bed material, and elutriated material have been proposed to improve SO_2 removal. A particle bonding device consists of a rotating drum or pan with a powder feed mechanism, a fog type water spray nozzle, and a plow. The drum or pan may be inclined at an angle such that the fully formed particles overflow after the appropriate residence time. The material to be processed in the particle bonding device is first pulverized into a powder. The powder is fed to the rotating pan or drum where nucleation occurs by the adhesion of several fine particles to a water droplet. The nucleated particle rolls due to the rotation of the equipment, picking up individual grains on its surface such that it grows in diameter. Growth continues with the large particles being buoyed up to the surface where overflow occurs. The plow prevents particles from attaching to the equipment surfaces. Particle bonding produces a uniform particle size with large macropores from the limestone, spent bed material, or elutriated material. The large macropores have high chemical reactivity due to their high surface-to-volume ratio.[23]

One conceptual design of a spent bed material particle bonding process for an AFBC boiler consists of the following steps:[24]

- Withdrawal and cooling the AFBC boiler spent bed material,
- Screen sizing the spent bed material,
- Milling the spent bed material,
- Blending the spent bed material with elutriated material removed from the flue gas,
- Particle bonding the elutriated/spent bed material blend,
- Steam curing the bonded particles, and
- Introducing the bonded particles to the boiler with the coal/limestone feed.

The overall limestone utilization is projected to improve due to reinjection of unreacted sorbent with a more reactive pore structure as a result of the particle bonding process. Limestone requirements are projected to be reduced by about 60 to 70 percent with the process.

The use of limestone utilization enhancement additives is currently being investigated as a means to improve SO_2 removal efficiency. The additives that have received the most attention are alkali salts such as NaCl, Na_2CO_3, Na_2SO_4, KCl, and $CaCl_2$. Salt addition to a limestone calcining environment results in formation of trace amounts of liquid on the calcined limestone particles, with subsequent recrystallization of the particles and reformation of the particles' pore structures.[25] Experimental programs have demonstrated up to two- or three-fold improvements in limestone's sulfation capacity with salt addition (usually NaCl or $CaCl_2$ ranging from 0.5 to 2.0 weight percent of the coal feed.[25,26,27] There are, however, potential corrosive effects associated with introducing certain salts to an AFBC boiler.

Grinding and reinjection of spent sorbent also has the potential to increase limestone utilization. Spent bed material typically contains a large fraction of unreacted sorbent. Much of the sorbent is, however, at the core of the particle and is isolated by a crust of calcium sulfate. Preliminary testing of grinding and reinjecting sorbent in an experimental AFBC unit resulted in an 18 percent improvement in SO_2 removal efficiency for a constant limestone feed rate.[28]

4.1.8 Demonstration of SO_2 Reduction

The current New Source Performance Standard (NSPS) for coal-fired boilers with heat inputs over 250×10^6 Btu/hr is 1.2 lb $SO_2/10^6$ Btu. Table 4.1-1 summarizes SO_2 emission control data along with the associated operating parameters for various industrial size AFBC boilers. The TVA 20 MWe unit with a solids recycle ratio of zero demonstrates the SO_2 removal achievable by conventional AFBC boilers without recycle. Solids recycle incorporated in the same unit is shown to provide a substantial improvement in sulfur capture.

For a comparison to first generation units, Georgetown University's AFBC boiler averages about 85 percent SO_2 removal with 3 percent sulfur coal at Ca/S ratios of between 3 and 6. However, significant design and operating problems have been encountered at this unit which have resulted in higher Ca/S ratios than originally anticipated.[29]

The staged bed units represented in the table were both designed by Wormser Engineering. The unit at the United Shoe Manufacturing Corporation (USMC) shows limited improvement in SO_2 removal compared to the conventional bubbling bed design without solids recycle. The Iowa Beef Processors (IBP) FBC boiler achieved lower SO_2 removal than the conventional design without solids recycle. However, the IBP data was taken from a test in which steady-state operation of the FBC was not achieved.

The Lurgi circulating bed data demonstrate a significant improvement in limestone utilization over the other design configurations and the ability

TABLE 4.1-1. SUMMARY OF INDUSTRIAL SIZE BOILER SO_2 EMISSIONS CONTROL DATA FROM SEVERAL AFBC CONFIGURATIONS

Configuration	Location	Heat Input 10^6 Btu/hr	Coal Sulfur Content Percent	Coal Sulfur Content lb SO_2/10^6 Btu	Type of Data[a]	Test Duration hrs	Ca/S Ratio	Recycle Ratio	Percent SO_2 Removal	Emissions lb/10^6 Btu
Conventional Bubbling Bed	TVA 20 MWe (9)	155	4.45	7.6	Cont 15	15	3.0	0	87	0.96
	TVA 20 MWe (9)	155	3.84	6.7	Cont 15	12	3.0	1.5	98	0.14
	Georgetown Univ. (29)	~120	1.7-2.7		Cont 15	30[b]	3-6	2	85	0.2-0.9
Staged Bed	United Shoe Manufacturing Corp. (18)	3	1.5	2.2	EPA M6	-	3.0	-	90	0.23
	Iowa Beef Processors (30)	88	4.21	6.7	EPA M6	9	3.0	-	82	1.19
Circulating Bed	Battelle MS-FBC (31)	50	1.5	2.30	-	-	4.5	-	95	-
	Lurgi (19)	-	-	-	-	-	1.5	-	90	-
	Plant A	54	2.0	4.18	-	-	3.5	-	90	0.42

[a] Cont 15: Continuous readings taken every 15 minutes.

EPA M6: EPA Method 6.

[b] Days

to achieve 90 percent SO_2 removal. The Battelle Multi-Solid Fluidized Bed Combustion (MS-FBC) data represent a very conservative design as indicated by early operation of the facility.

Contacts were made with eight operators of coal-fired AFBC boilers. Of the seven responses received to date, only three units (Plants A, B, and C in Table 3.2-8) are in operation and using limestone to control SO_2 emissions. Plant A features a circulating bed design and achieves 90 percent SO_2 removal with a Ca/S ratio of 3.5 (refer to Table 3.2-8 for other operating parameters). Although Plants B and C are operational, SO_2 emissions data are not available.

4.2 SUMMARY OF NO_x EMISSION DATA

While the potential for reducing NO_x emissions from AFBC units has been recognized in the past, the major emphasis has been on optimizing combustion efficiency and SO_2 control. As a result, most test data do not reflect emissions at conditions selected to optimize NO_x control. However, recent test data more clearly illustrate the capability of AFBC systems to reduce NO_x emissions.

The ITAR identified the following design and operating factors for conventional bubbling bed AFBC systems which influence the formation and control of NO_x:[1]

- Bed temperature,
- Fuel nitrogen,
- Coal particle size,
- Excess air,
- Gas residence time (bed depth and superficial gas velocity), and

- Factors affecting localized reducing reaction conditions in the system.

Each of these variables affects NO_x emissions and will be briefly reviewed. Data on the effects of solids recycle, staged combustion air, staged beds, and circulating beds will be presented to demonstrate the advantages of these more recent design configurations with regard to NO_x emissions. A summary of the NO_x emissions data for the different design configurations will also be presented.

4.2.1 Design Variables Affecting NO_x Emissions

One of the advantages of AFBC over conventional coal combustion methods is the low level of NO_x emissions produced due to the lower combustion temperatures. Normal AFBC operating temperatures are in the range of 1400° to 1650°F. The ITAR identified research which observed an increase in NO_x emissions with increasing bed temperatures up to approximately 1450° to 1550°F.[32] Above this temperature, NO_x emissions were observed to decrease slightly. Above 1650° to 1830°F, thermal NO_x formation became significant, and the emission rate of NO_x began to increase.

Since the low combustion temperature in an AFBC boiler significantly suppresses the thermal fixation of atmospheric nitrogen, NO_x emissions primarily result from the conversion of fuel nitrogen. The ITAR identified research which attributed 90 percent of the NO_x emissions to nitrogen compounds in the fuel, with only 10 percent due to the fixation of atmospheric nitrogen.[33]

In addition to coal nitrogen content, the ITAR identified research which investigated the effect of coal particle size on NO_x emissions, although the results of the research are conflicting.[34,35] Recent research has determined that coal size is of minor importance when compared to other design variables such as bed temperature and excess air ratio.[10]

Most experimental data on the effect of excess air on NO_x emissions have been measured at air stoichiometries of from 0.9 to 1.2. In this range, NO_x emissions rise sharply as the air flow is increased. This rise in NO_x emissions is apparently related to a large decrease in CO available for NO_x reduction reactions as air rates rise to and above stoichiometric levels. Above a stoichiometric air rate of 1.2, further increases in air rate have a much smaller effect on NO_x emissions. Also, decreases below a stoichiometric ratio of 0.9 have been shown to have limited effect on NO_x emissions.[10,36,37] Figure 4.2-1 demonstrates the effect of stoichiometric air ratio on NO_x emissions for a conventional AFBC boiler.[10]

Another factor which affects NO_x emissions from an AFBC boiler is the gas phase residence time, defined as the ratio of expanded bed depth to superficial gas velocity. The ITAR recognized the inverse relationship between NO_x emissions and gas phase residence time. Longer residence time in the fuel zone increases the rate of the reducing reaction between NO_x and char or CO resulting in lower NO_x emissions. Recent research data, presented in Figure 4.2-2, illustrate the effect of gas phase residence time on NO_x emissions.[12]

The ITAR identified several factors which affect the local reducing conditions responsible for the conversion of NO_x to elemental nitrogen. Among these are gas phase residence time and bed temperature which have previously been reviewed. In addition, volatile coal constituents, especially ammonia, and $CaSO_3$ may react with NO to produce elemental nitrogen. The postulated reactions are presented in the ITAR.

4.2.2 Solids Recycle

Recycle of elutriated solids decreases NO_x emissions and increases SO_2 removal and combustion efficiency. The TVA 20 MWe pilot plant produced NO_x emissions from 0.29 to 0.40 $lb/10^6$ Btu for operation without recycle of elutriated material. A solids recycle ratio ranging from 1 to 3 lowered the NO_x emissions to ranges of 0.19 to 0.26 $lb/10^6$ Btu.[9] Apparently, carbon in the recycled elutriated solids is available for heterogeneous reduction reactions between NO and carbon.[38,39,40,41]

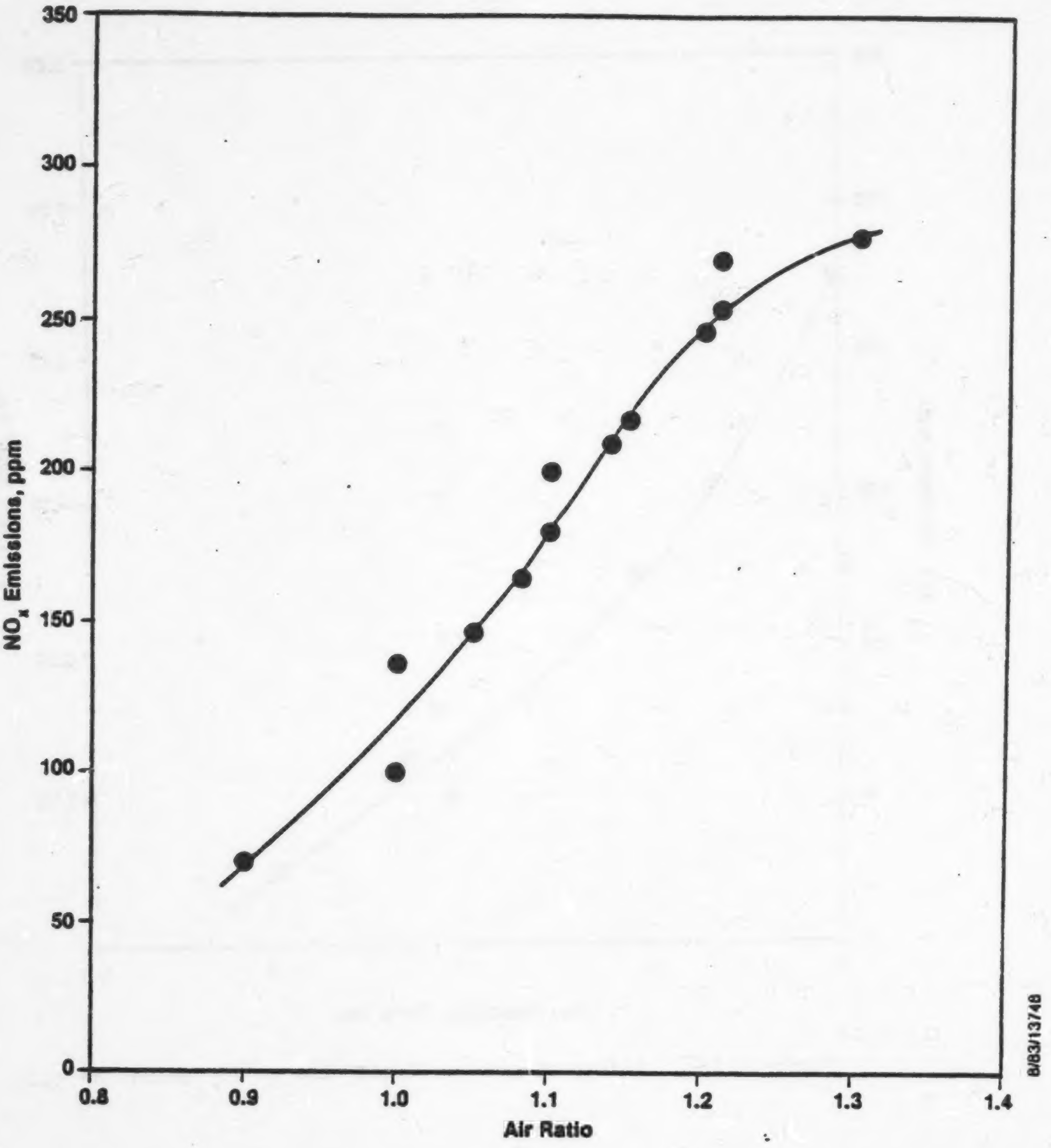

Figure 4.2-1. Effect of stoichiometric air ratio on NO_x emissions in a conventional AFBC (10)

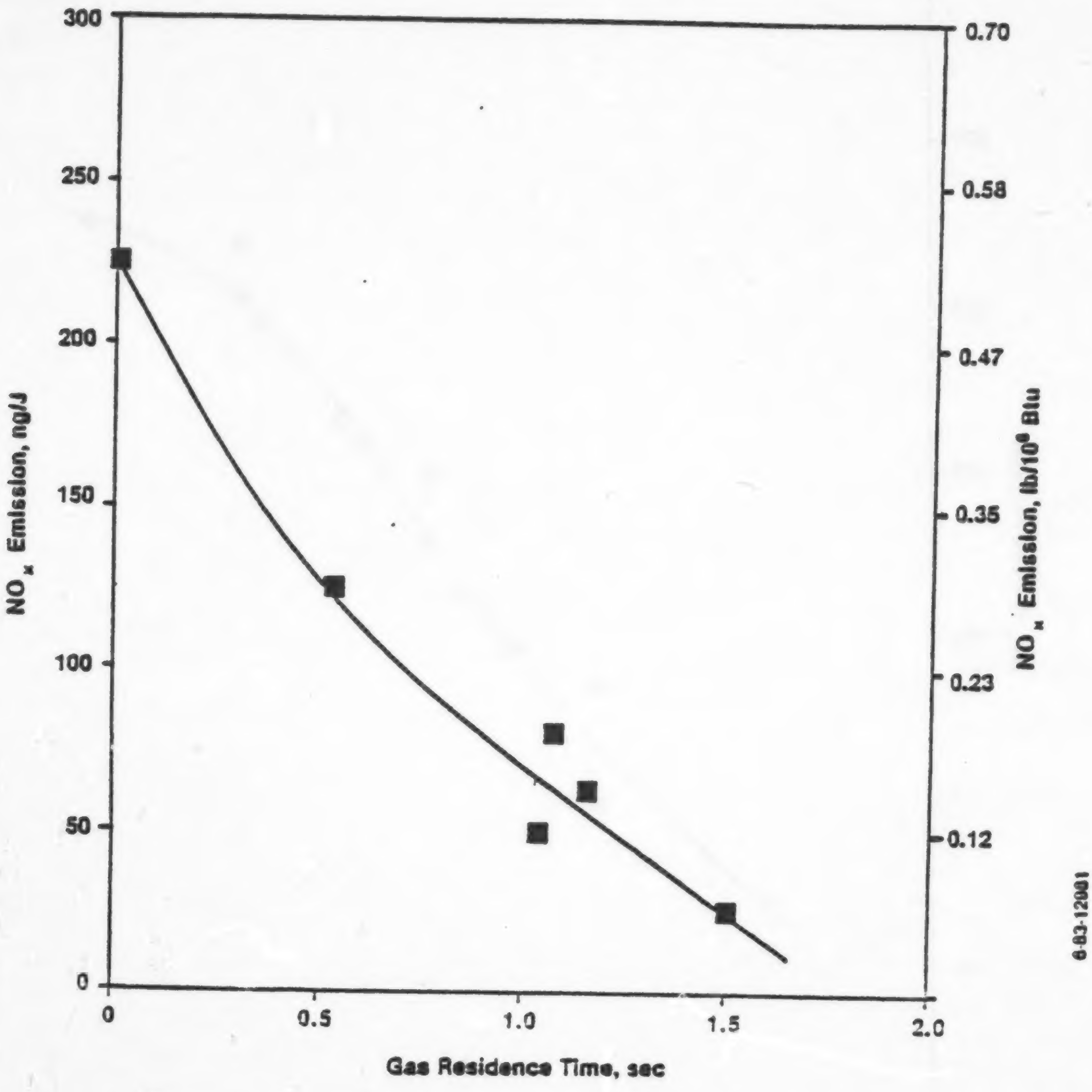

Figure 4.2-2. Effect of gas residence time on NO$_x$ emissions (12)

4.2.3 Staged Combustion Air

Early testing of staged combustion air demonstrated its ability to reduce NO_x emissions by up to 50 percent.[36] Tests conducted at the EPRI/B&W 6'x6' unit show that NO_x emissions resulting from the use of staged air can be reduced to 0.15 $lb/10^6$ Btu from 0.5 $lb/10^6$ Btu without staged air.[42]

The variable with the greatest impact on NO_x emissions for staged combustion air is the primary/stoichiometric air ratio, defined as the ratio of air introduced through the distributor plate to the calculated stoichiometric air. Figure 4.2-3 illustrates the effect of this air ratio on NO_x emissions from a Battelle test unit. Operation of an AFBC boiler with primary/stoichiometric air ratios less than 1.0 results in the creation of a reducing zone. This promotes the reduction of NO by char and carbon monoxide.[10]

As stated previously, a tradeoff exists between NO_x and SO_2 emissions when the combustion air is staged for NO_x control. (Refer to Section 4.3 for a discussion of this tradeoff.)

4.2.4 Staged Beds

Staged bed AFBC boilers are designed to achieve low NO_x emissions by operating with the lower bed at substoichiometric conditions; the balance of the air necessary for combustion is added in the second bed. The only steady-state data available for this configuration are from the United Shoe Manufacturing Corporation's (USMC) Wormser unit. Emissions of NO_x averaged 0.35 $lb/10^6$ Btu which is above the NO_x emission level achievable by a conventional bubbling bed AFBC without solids recycle.[43] Short-term testing of a Wormser unit at Iowa Beef Processors in March, 1983 demonstrated NO_x emissions generally between 0.25 and 0.55 $lb/10^6$ Btu, but operating conditions were fluctuating.[44]

4.2.5 Circulating Bed

Circulating bed AFBC boilers feature very extensive recirculation of elutriated solids. In addition, staged combustion is often employed. Both of these techniques have been previously described as being effective for

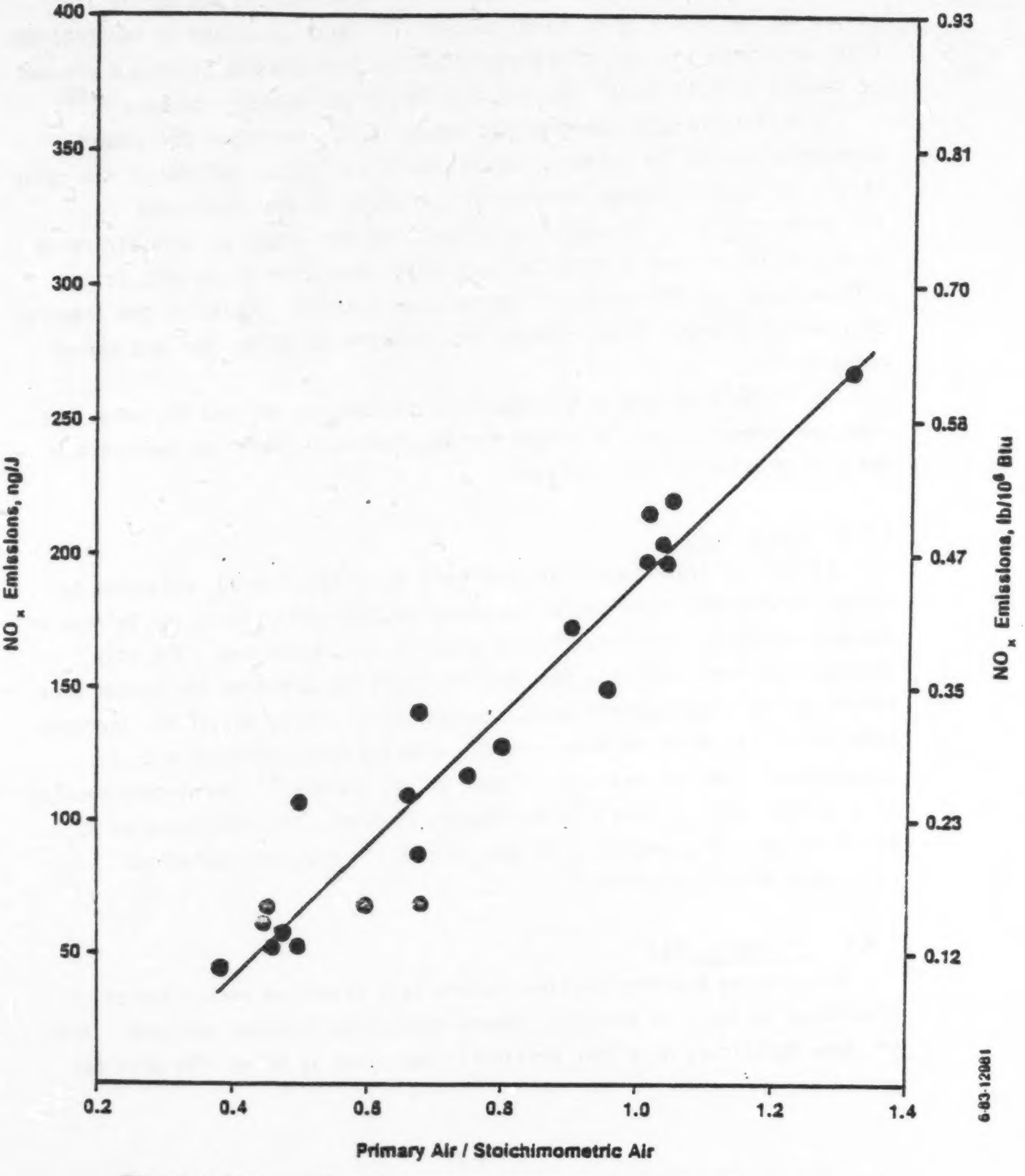

Figure 4.2-3. Effect of primary air/stoichiometric air ratio on NO$_x$ emissions (12)

reducing NO_x emissions. Figure 4.2-4 demonstrates the NO_x emissions from the Battelle $1x10^6$ Btu/hr test unit with staged combustion air.[20] The lowest NO_x emission level achieved, 0.15 $lb/10^6$ Btu, was with a primary/stoichiometric air ratio of 0.5.

4.2.6 Demonstration of NO_x Reduction

Table 4.2-1 summarizes NO_x emissions data for the newer AFBC design configurations. For comparison with first generation AFBC boilers, the Georgetown University unit averages about 0.50 $lb/10^6$ Btu.[45] The effect of solids recycle on NO_x emissions for conventional bubbling beds is illustrated by data from TVA's 20 MWe pilot plant. In addition, the table shows that staged combustion air significantly decreased NO_x emissions at B&W's 6'x6' test unit. The NO_x emissions control achievable by circulating bed AFBC boilers with staged combustion air is illustrated by data from the Battelle MS-FBC process. The NO_x emissions data from Wormser's staged bed process are also presented.

Several points should be emphasized when examining the results in Table 4.2-1. First, long-term testing at conditions producing very low NO_x emissions, especially substoichiometric firing, has not been conducted. Also, issues concerning proper materials of construction in reducing regions in the unit have not been resolved. Finally, the data presented for NO_x and SO_2 emissions do not necessarily reflect emissions control that can be obtained simultaneously. While the interactions between SO_2 and NO_x emissions must be further defined to establish optimum performance, the trends in Tables 4.1-1 and 4.2-1 illustrate that factors such as solids recycle, staged beds, and circulating bed designs can be used to reduce both SO_2 and NO_x emissions.

4.3 SO_2/NO_x TRADEOFF

Most design and operating factors which affect both SO_2 and NO_x can be set to simultaneously reduce NO_x and SO_2 emissions. These factors include bed temperature, gas residence time, and solids recycle. However, the

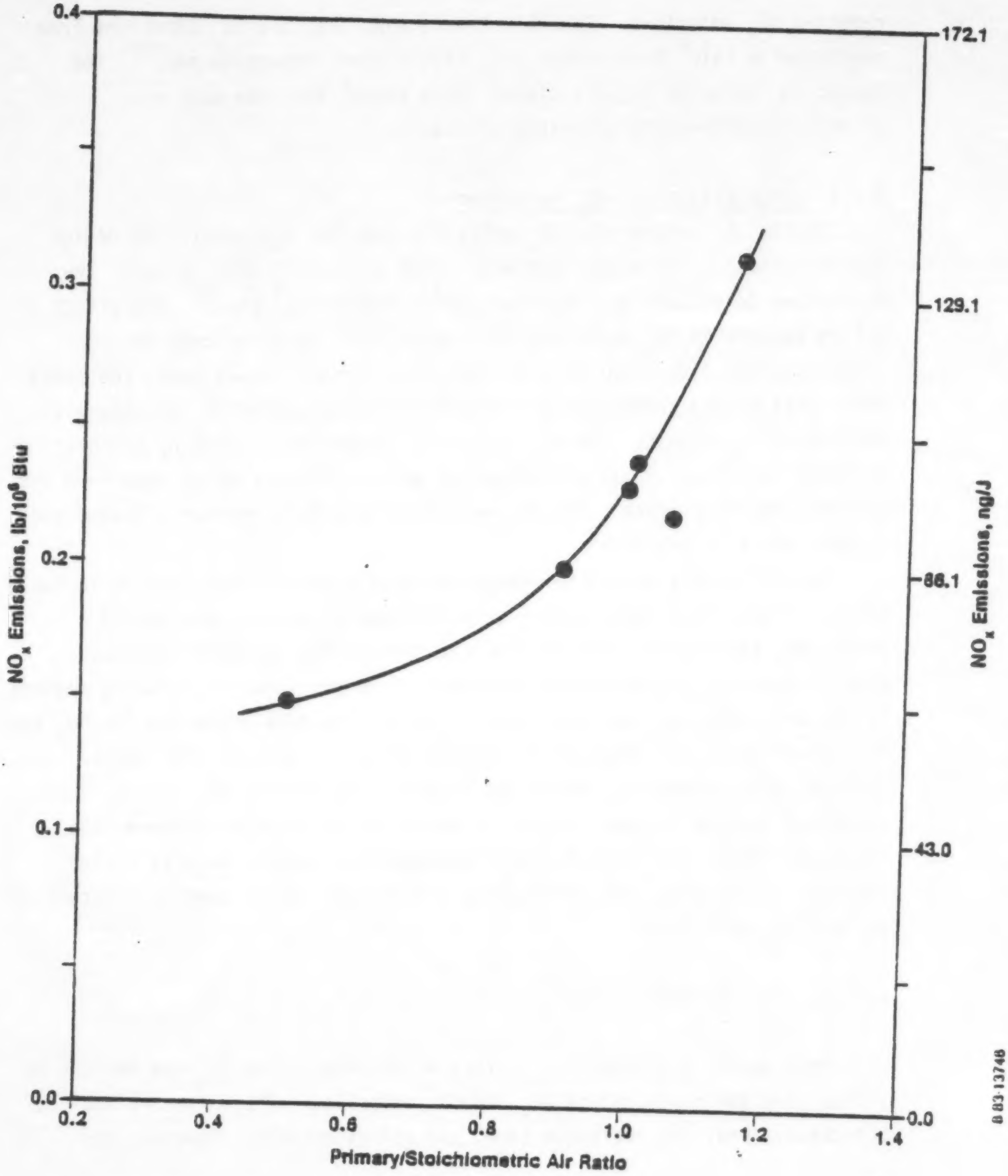

Figure 4.2-4. NO$_x$ emission test results for circulating bed AFBC with staged combustion air (20)

TABLE 4.2-1. SUMMARY OF NO_x EMISSIONS FOR VARIOUS AFBC CONFIGURATIONS

Configuration	Location	Heat Input 10^6 Btu/hr	Type of Data[a]	Test Duration, hrs	Primary/Stoich. Air Ratio	NO_x Emissions $lb/10^6$ Btu	Recycle Ratio
Conventional Bubbling Bed	TVA 20 MWe (9)	155	Cont 15	15	-	0.34	0
	TVA 20 MWe (9)	155	Cont 15	12	-	0.23	1-3
	B&W 6'x6' (42)	24	-	-	-	0.15	0
Staged Bed	United Shoe Manufacturing Corp. (43)	3	-	-	-	0.35	-
Circulating Bed	Battelle MS-FBC (20)	1	-	-	0.50	0.15	-
		1	-	-	0.90	0.20	-
		1	-	-	1.15	0.33	-

[a] Continuous readings were taken every 15 minutes.

primary operating conditions used to reduce NO_x emissions, low excess air and staged combustion air, involve a tradeoff with SO_2 emissions. Low excess air and staged combustion air were shown in Figures 4.2-1 and 4.2-3, respectively, to decrease NO_x emissions. However, these NO_x emission reduction methods were shown in Section 4.1.1 and Figures 4.1-5 and 4.1-6 to increase SO_2 emissions. Staged combustion air test results, in which both SO_2 and NO_x emissions were measured, are presented in Figure 4.3-1.[46] As the primary air ratio was lowered from 1.04 to 0.87, NO_x emissions dropped from 240 to 90 ppm, and SO_2 removal decreased from 95 to 90 percent. The increase in SO_2 emissions is small compared to the reduction in NO_x emissions and can be offset by increasing the Ca/S ratio and/or the solids recycle ratio. It should be noted, however, that both of these methods involve an increase in operating costs.

4.4 PARTICULATE MATTER EMISSION DATA

The following design factors were identified by the ITAR as being important to the quantity of particulate matter (PM) emitted from an AFBC boiler:

- Coal
    - ash content
    - sulfur content
    - agglomeration characteristics

- Sorbent
    - particle size
    - attrition and decrepitation characteristics

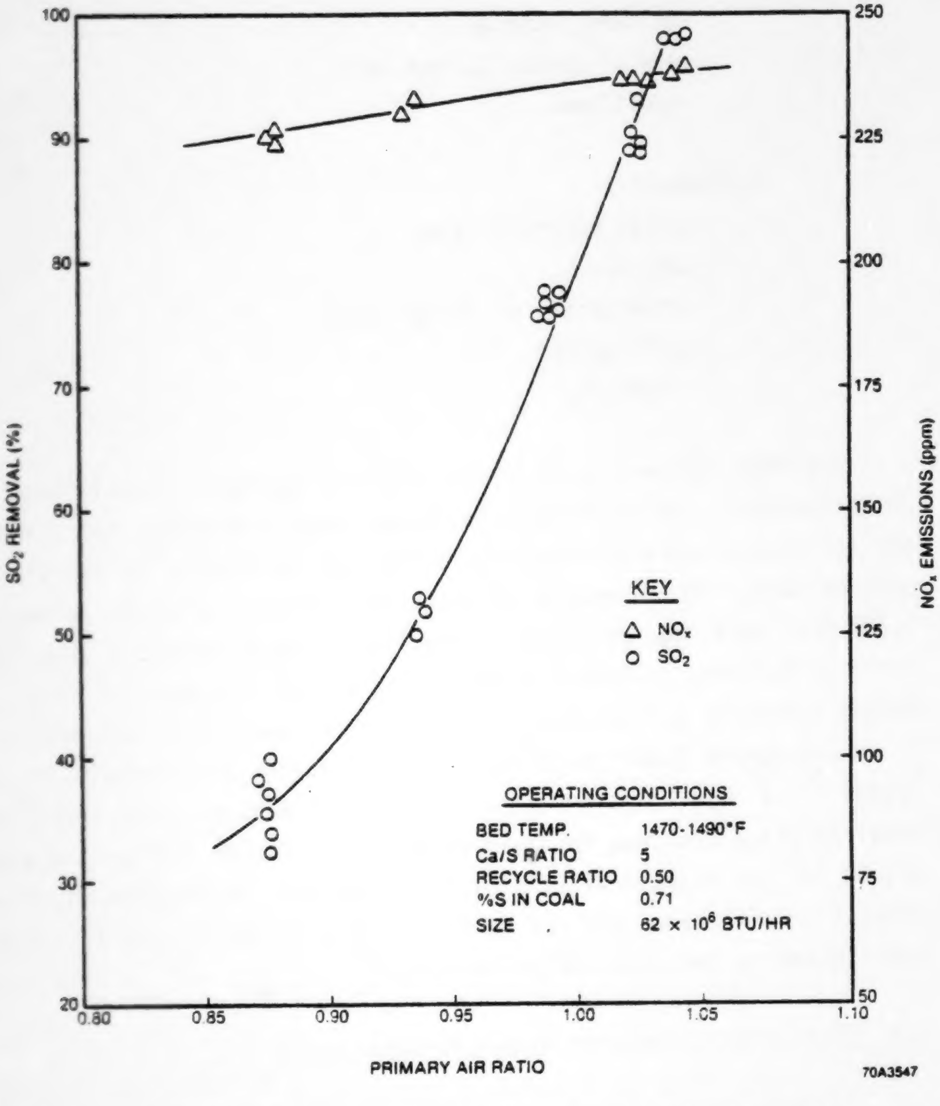

Figure 4.3-1. NO_x/SO_2 Tradeoff for Staged Combustion Air.

- Operation
    - -superficial velocity
    - -primary recycle
    - -use of carbon burnup cell
    - -additives

- Bed Geometry
    - -cross sectional area
    - -bed depth
    - -orientation of boiler tubes
    - -grid design
    - -freeboard

Cyclones followed by a fabric filter or an ESP have both been used for PM collection. Fabric filters have been used more widely for commercial applications instead of ESPs due to the low resistivity of ash produced by AFBC boilers. PM collection efficiencies of 99.81 to 99.94 percent (<0.03 lb/MM Btu) have been obtained at the TVA 20 MWe pilot plant with the use of cyclones followed by fabric filters with a 1.48 air-to-cloth ratio.[9] EPA Method 5 testing for particulate emissions at Georgetown University resulted in an average of 0.065 lb/10^6 Btu for the cyclone and baghouse PM collection system.[29] A PM collection efficiency of 99.7 percent (0.06 lb/10^6 Btu) was obtained using cyclones followed by an ESP (effective collection area of 21,000 ft^2) at a paper mill in Kauttua, Finland. A consistently high combustion efficiency and low carbon content in the fly ash may have contributed to the good ESP performance.[47]

4.5 OTHER FACTORS RELATED TO BOILER PERFORMANCE

As indicated in the ITAR and in the preceding discussions, considerable research emphasis has been directed towards the environmental characterization of FBC technology. Furthermore, significant development work has been undertaken to improve the environmental performance of the technology. However, other technical issues which are important to the development of AFBC boiler technology for industrial boiler use have received recent attention. These include:

- Boiler efficiency,
- Solid waste impacts,
- Fuel use flexibility,
- Erosion/corrosion, and
- Turndown characteristics.

These performance factors and their relation to recent improvements in FBC technology are reviewed in this section.

4.5.1 Boiler Efficiency

Boiler efficiency is defined as the percentage of the total energy (fuel) input that is available for the generation of steam. Conventional coal-fired industrial boilers typically achieve boiler efficiencies ranging from approximately 80 percent to 85 percent, depending on design configuration and coal type. By comparison, recent demonstration plant testing of state-of-the-art bubbling bed FBC technology has also shown boiler efficiency values of 80 to 85 percent.[9] The portion of the total energy input that is not available for steam production consists of (1) flue gas heat losses, (2) hot solids heat losses, (3) net calcination and sulfation reaction heat losses, (4) unburned carbon heat losses, and (5) radiation and miscellaneous heat losses.[48]

Flue gas heat losses (in the form of sensible heat and the latent heat of water vaporization) represent the major heat loss from industrial boilers, typically approximating 10 to 15 percent of the total fuel energy input. Traditional and advanced FBC boiler designs tend to have lower flue gas heat losses than conventional coal-fired industrial boilers primarily because of lower excess air rates. FBC technologies typically feature excess air rates of about 20 percent compared to levels as high as 50 percent for industrial spreader stoker boilers.

Also, the lower excess air levels and increased heat transfer rates of FBC designs due to turbulent and well-mixed combustion zones allow for more compact boiler designs. It is expected that, as the technology matures, shop-fabricated package FBC boilers will be commercially available in steam generation capacities greater than those available for conventional coal-fired package boilers (currently about 200 x 10^6 Btu/hr).

Heat losses due to hot solids generation (spent sorbent products and bottom and fly ash) are typically somewhat greater for traditional and advanced FBC configurations than for conventional coal-fired boilers. This result is due to the presence of increased solids levels, i.e., in-situ sorbent products, in FBC boilers. Development work aimed at minimization of solids heat losses has focused on reduction of Ca/S ratio and heat recovery from spent bed material.

Net heat losses (or gains) due to calcination and sulfation reactions in the boiler are inherent to FBC operation. Calcination and sulfation reactions are endothermic and exothermic, respectively, and their heat effects are off-setting. Depending on the Ca/S ratio, sorbent utilization rate, and SO_2 emission limits, the net effect may be a heat loss or a heat gain.

Unburned carbon heat losses are typically expressed in terms of combustion efficiency. Development efforts have targeted combustion efficiency levels at 95 to 99 percent for FBC boiler technology so that it can compete with conventional coal combustion in this area. First generation FBC boilers often failed to meet the targeted combustion efficiency level, even with a carbon burn-up cell or solids recycle. However, recent improvements in AFBC-with-recycle operation and development of novel configurations, e.g., circulating fluidized beds, have enabled 95 to 99 percent combustion efficiency levels to be achieved.

Radiation and miscellaneous boiler heat losses, typically a minor component of the total heat losses, are not expected to differ significantly for FBC as compared to conventional coal combustion technology. However, FBC technology may have the potential for somewhat lower radiation losses due to lower operating temperatures and more compact boiler designs.

4.5.2 Solid Waste Impacts

Solid waste from FBC boilers differs in composition from that produced in conventional coal-fired boilers. FBC waste typically contains greater amounts of carbon, calcium, and sulfur-bearing compounds. The amount of solids from an FBC boiler is expected to equal or exceed those from a conventional coal-fired boiler with FGD.[37] The amount of solids generated in an FBC boiler is a function of (1) unit size, or coal feed rate, (2) Ca/S ratio, or sorbent feed rate, (3) coal and sorbent properties, (4) coal combustion efficiency, (5) degree of sorbent utilization, (6) SO_2 and particulate emission levels, and (7) unit configuration (e.g., AFBC or PFBC).

Two options are available to the industrial AFBC boiler user with respect to alleviating solid waste impacts. These options are to market the solid waste as a useful by-product or to dispose of the waste in an environmentally acceptable manner.

The marketing option is currently less feasible than the disposal option for the potential industrial AFBC user. Potential markets for AFBC solid waste appear to be competitive and limited (e.g., construction materials market) or undefined (e.g., agricultural supplements market). Unresolved questions remain regarding the technical feasibility and environmental acceptability of converting FBC solid wastes into useful resources. Applications that have received considerable research emphasis include the use of FBC solid waste as construction material additives, agricultural supplements, acidic waste treatment agents, and road base material.[48,49,50,51,52,53,54]

Because of apparently limited market potential for FBC solid wastes, most FBC waste generated in the near-term will have to be disposed of in a manner consistent with applicable regulations. It appears that the most significant regulations regarding disposal, in terms of cost to the AFBC boiler use, are those associated with the Resource Conservation and Recovery Act (RCRA).[37]

Hazardous characteristics currently defined by RCRA provisions are ignitability, corrosivity, reactivity, and toxicity. The only characteristic that may be applicable to FBC waste appears to be toxicity; however, laboratory studies have indicated that typical FBC wastes would not be classified as hazardous according to toxicity characteristics.[55,56] Of course, toxicity characteristics of FBC waste (and, ultimately, RCRA classification as hazardous or nonhazardous) are dependent on specific coal and sorbent properties, so additional data are necessary to conclusively evaluate the classification of AFBC solid waste.

Recent data suggest that FBC solid waste can satisfy the RCRA requirements for sanitary landfill disposal, i.e., ground water at the disposal site boundary should be able to satisfy the National Interim Primary Drinking Water Regulations (NIPDWR).[37] Nonetheless, potential environmental problems of landfilling remain, including (1) heat release from the solid waste as CaO hydrates to $Ca(OH)_2$ upon exposure to moisture, and (2) leachate characteristics, especially excessive pH, total dissolved solids (TDS) content, and sulfate content.[47]

Recent improvements in design configuration, including recycle and circulating bed options, have served to lessen the amounts of solid waste generated, primarily through the use of lower Ca/S ratios.

4.5.3 Fuel Use Flexibility

A significant advantage of FBC technology that has spurred its development is its ability to efficiently burn a wide variety of fuels. A given FBC boiler design will not necessarily burn any type of fuel; nonetheless, a specific unit can handle considerably wider fluctuations in fuel composition than a conventional combustion boiler. Recent design developments, such as the circulating bed principle, have further enhanced AFBC fuel flexibility.

The focus of the discussions presented in the ITAR and in this document has been on FBC firing of coal. However, FBC technology has been shown to satisfactorily burn a wide variety of fuels, including coal processing wastes, oil shale, petroleum coke, waste wood, municipal waste, dried sewage sludge, and other agricultural and industrial wastes.[31,57,58,59] Several investigations of alternate fuel feasibility have been performed at the pilot or demonstration scale, e.g., the Shamokin anthracite culm project.[57] However, alternate or low-grade fuels have also been fired in commercial installations (e.g., Conoco's South Texas circulating bed design firing coal and petroleum coke and over 2000 AFBC units firing low-grade coals and industrial wastes in the People's Republic of China.[31,60]

4.5.4 Erosion/Corrosion

A significant amount of research has been undertaken to identify the erosion/corrosion parameters and the potential for various FBC design configurations. Earlier theories maintaining that corrosion in traditional bubbling beds would not be significant because of the low-temperature operation of the combustion zone have been rejected. Recent research has shown that sulfidation/oxidation of metallic components does occur in FBC bubbling beds, and that selection of tube material is critical in control of these corrosion mechanisms.[61] AFBC units which operate under substoichiometric conditions to reduce NO_x formations also have potential corrosion problems due to the reducing environment.

The potential for erosion of boiler internals is enhanced by circulating fluidized bed technology, due to impingement of high-velocity particles on interior boiler surfaces. However, the potential for tube corrosion is reduced because heat transfer surface is less likely to be located in a reducing zone.[48] Conversely, staged air and staged bed configurations, by the nature of their design and operation, include reducing zones in their combustion regions. This feature enhances the possibility of metal corrosion; as result, heat transfer surface is either excluded from these zones or is made of an appropriate alloy metal.

Erosion/corrosion issues have been a major impediment in the commerical development of PFBC technology. Significant research activity has been undertaken to resolve problems associated with corrosion and erosion of components of gas turbines powered by PFBC exhaust gases.[62,63] Continuing activity in this area is necessary to bring PFBC technology closer to commercialization.

4.5.5 Turndown

A major technical problem associated with first generation traditional AFBC designs was load turndown. The following methods were initially used to control the amount of heat transferred to the boiler tubes: (1) bed segment slumping, (2) temperature variation, and (3) bed height variation. Problems were encountered with these methods, including failure to refluidize slumped portions of the bed, compromise of SO_2 reduction performance due to temperature swings, and difficulty in controlling bed height to the desired level. Early target turndown levels for industrial AFBC boilers approximated a ratio of 4:1. Newer design configurations have incorporated improvements with regard to load turndown. The implementation of solids recycle has provided more flexibility in load control for bubbling bed designs. The recycle solids flow rate provides an additional parameter that can be varied to effect changes in heat transfer rate. Similarly, the circulating bed designs feature load control by variation of the solids recirculation rate. Finally, the separation of combustion and desulfurization reactions in the staged bed designs permits greater flexibility with regard to load control. These features have allowed the current turndown ratio of 4:1 to be achieved. However, it should be noted that turndown is very complicated and can significantly effect emissions and overall AFBC performance.

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SECTION 5
FBC COST ALGORITHM DEVELOPMENT

Cost algorithms are used in this study to estimate capital and operating costs for FBC systems, as well as conventional boilers, over a wide range of system sizes and operating conditions. An algorithm is a mathematical expression which relates costs to key design and operating parameters (e.g., boiler size, coal properties, raw material costs). One advantage to the use of algorithms is that they can be loaded onto a computer to allow efficient cost estimating for a large number of cases.

Cost algorithms have already been developed for both conventional boilers and FGD systems and are well documented in other reports.[1,2] A major objective of this study has been to develop a workable, up-to-date, and valid cost algorithm for industrial-size FBC systems. The development of the FBC algorithm is described in this chapter as well as validation of the algorithm with vendor-developed cost estimates.

5.1 BASIS OF DESIGN

The discussion in Sections 3 and 4 makes the point that three major FBC boiler design types are being offered on a commercial basis to buyers in the industrial boiler market: conventional "bubbling" FBC boilers, circulating FBC boilers, and two-stage FBC boilers. Pressurized FBC technology is in a relatively early stage of development and is more suitable for utility applications than industrial steam generation. Although the circulating and two-stage FBC boiler designs are making significant inroads in the industrial sector, the information in Tables 3.2-6 to 3.2-8 indicates that the majority of existing and planned FBC units are of the conventional bubbling bed design. Given the conservative nature of the industrial boiler market and the fact that circulating and two-stage FBC boilers are in an earlier commercialization stage than conventional FBC boilers, it is likely that a great majority of the industrial FBC systems installed over the next five years will be atmospheric, conventional FBC units. Accordingly, the conventional AFBC boiler design has been chosen as the basis of the FBC algorithm.

It is of interest to note, however, that the limited amount of cost data available comparing atmospheric, circulating FBC to conventional FBC indicate that CFBC capital costs are similar to those of conventional FBC systems, while operating costs for CFBC are estimated to be slightly less.[10] A 1979 cost comparison of both systems in an industrial setting (meeting a 1.8 lb $SO_2/10^6$ Btu limit) found both the capital and operating costs of the systems to be within the accuracy range (± 25 percent) of the study.

5.1.1 Comparison of Design Bases

One of the most extensive set of analyses currently available which relates FBC design and operating factors to SO_2, NO_x, and PM emissions is contained in the FBC ITAR. Much of that discussion has been summarized in Sections 3 and 4 of this report. The ITAR analyses assumes that the "best system" of SO_2 emissions reduction is one which minimizes sorbent feed rates and still attains high levels of emissions control. The experimental results and theoretical considerations discussed in the ITAR indicate that "small particle sizes (in the range of 500 μm) and sufficiently long gas phase residence time (0.67 sec.) are representative conditions for effective SO_2 control, although most FBC facilities currently are designed or operated with shorter residence times and coarser particles."[3] The conditions specified in the ITAR for this "best system" of SO_2 control are listed in the first column of Table 5.1-1.

Because of the depth of analyses and consideration of emission and cost impacts which support this design basis, this basis been used for the purposes of algorithm development. A more pragmatic consideration is that an existing FBC cost algorithm has already been developed on the basis of this "best system" design. Thus only a review of the existing algorithm, and possibly minor modifications, are to provide a suitable algorithm for the purposes of this report.

The ITAR "best system" design basis was formulated from information and data available in the 1978-1979 time frame. Before accepting this design

TABLE 5.1-1. AFBC DESIGN/OPERATING CONDITIONS FOR THE ITAR MODEL PLANT AND THE TVA AND GU FACILITIES

	ITAR[a] "Best System"	TVA[b] Campaign I	GU[c] 1982 Tests
Design Basis Variables			
Bed Dept, ft	4	3.75	4.5
Superficial Gas Vel., ft/sec	6	9	8
Residence Time, sec	0.67	0.42	0.56
In-Bed Sorbent Part. Size, μm	600 - 700[d]	1,086[e]	≥1,000
Coal/Sorbent Feed System	Inbed/Overbed	Inbed	Overbed
Solids Recycle Ratio	0.2 - 0.4	0 - 1.5	2.2
Bed Temperature, °F	1,550	1,530	1,590
Excess Air, percent	20	22	20
Boiler Efficiency, percent	79 - 85	75 - 85	~80
Algorithm Input Variables			
Sorbent Reactivity	Medium	Medium	Low
SO_2 Removal, percent	90	87 - 98[g]	80 - 95
Ca/S Ratio	3.3	3.0	3 - 7
Coal Type	Eastern Bituminous	Eastern Bituminous	Eastern Bituminous
Coal Sulfur, percent	3.5	4.2	1.7 - 3.5
Coal Heating Value, Btu/lb	11,800	~12,000	+12,000
Heat Input, 10^6 Btu/hr	30 - 200	~165	~120

[a] Source: Reference 3.

[b] Source: Reference 5.

[c] Source: References 3 and 4.

[d] 600 to 700 μm mass mean particle size is equivalent to 500 μm surface mean particle size.

[e] Geometric mass mean particle size of bed drain material

[f] Estimate based on actual PM emissions and assumed cyclone efficiency of 90 percent.

[g] Higher freeboard may have contributed to higher SO2 removal values.

basis as representative of currently available technology, it is useful to compare it with the design bases of existing operating systems. Two such systems are the TVA 20 MW_e AFBC pilot plant and the Georgetown University (GU) FBC industrial boiler. These plants are generally representative of AFBC systems being offered commercially to industrial plant owners.

The second column in Table 5.1-1 lists the conditions of the TVA pilot plant during Campaign I testing. The final column summarizes the operating conditions for the GU boiler which are representative of the conditions in effect during the January/February 1982 emissions test series sponsored by EPA.[4]

The table shows that the design bases for these large, operating systems are comparable to the "best system" conditions of the ITAR, upon which the ITAR cost estimates, and ultimately, the FBC cost algorithm, are based. This comparison demonstrates that the design/operating conditions for industrial FBC units installed today, or in the next five years, will not be fundamentally different from the ITAR design basis. The fact that the gas residence time for the ITAR system is less than that for industrial installations suggests that ITAR estimates of boiler costs may be slightly higher than those for operating units.

5.1.2 Selection of Ca/S Ratios

One of the most important of the Table 5.1-1 parameters from the standpoint of SO_2 control is the Ca/S ratio. The data and discussion of Sections 3 and 4 and the FBC ITAR show that, for a given target SO_2 removal level, the Ca/S ratio in a conventional AFBC unit is primarily a function coal type, bed temperature, recycle ratio, sorbent reactivity, sorbent particle size, and gas residence time in the fluidized bed. The Ca/S ratios specified in the ITAR are based on experimental data collected on bench- and pilot-scale units operating over a wide range of conditions. The Ca/S ratios plotted in Figure 5.1-1 correspond to these data plus "best system" design/operating conditions. Also plotted on the same figure are performance data from the Georgetown University, B & W 6'x6', and TVA facilities. These units have been selected for comparison because they are

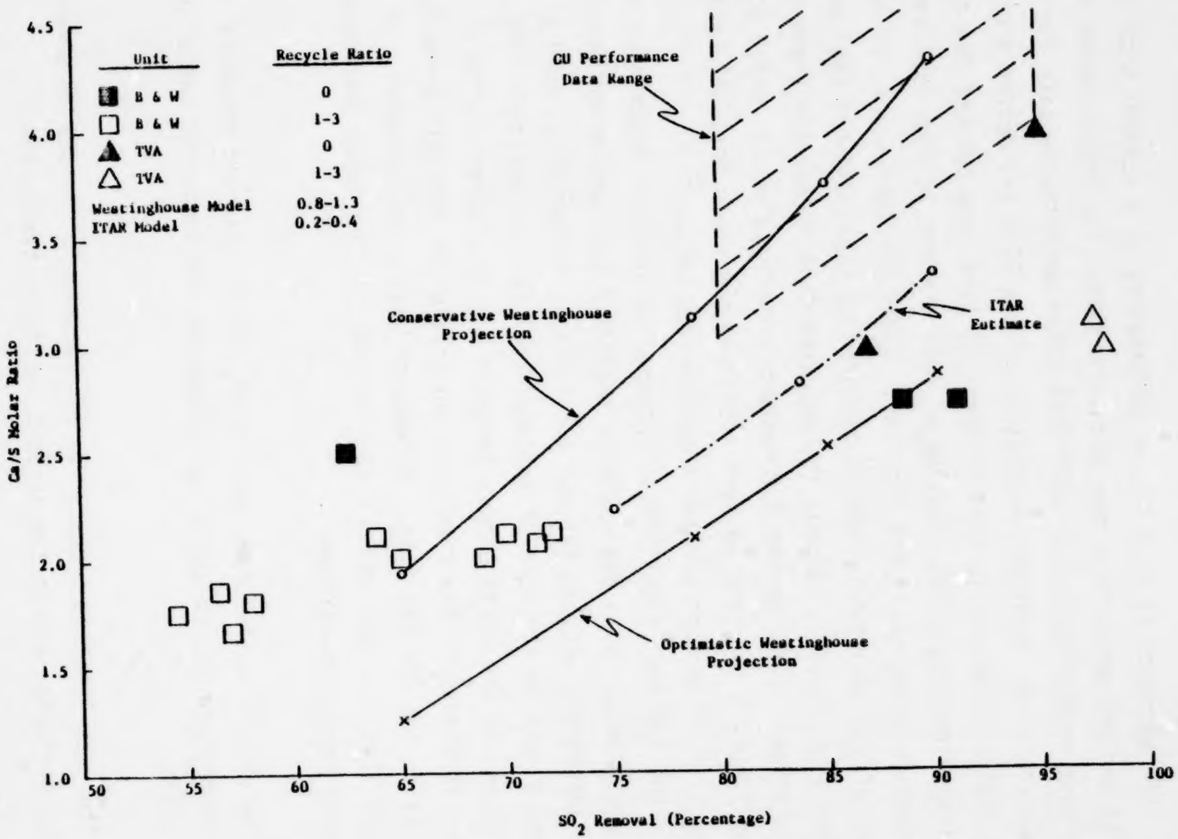

Figure 5.1-1. Ca/S Versus SO_2 Removal For Industrial AFBC Facilities Operating on High Sulfur Eastern Coal.

of a scale similar to commercial industrial FBC systems of conventional bed design.

The ITAR estimate in this figure corresponds to a sorbent with medium reactivity and 500 μm surface mean particle size. The figure shows that the ITAR estimate agrees reasonably well with other performance data for eastern bituminous coal. An important limitation of the ITAR estimation procedure for Ca/S ratios, however, is that it does not take into account the impact of alkali species (e.g., CaO, MgO, Na_2O, K_2O) present in some coal ashes, notably subbituminous coals and lignites. Under FBC conditions, as much as 50 percent of the coal sulfur can be captured by subbituminous coal ash. This effect significantly reduces the required Ca/S ratios for these coals. While this effect is not marked for eastern bituminous coals, which are the subject of Figure 5.1-1, for western subbituminous coals the ITAR Ca/S ratios are over 70 percent greater than reported values.[6]

Since the FBC cost algorithm is intended fo use with bituminous and subbituminous coals, it is desirable to include a Ca/S estimated methodology that will adequately account for ash alkalinity. Fortunately, such a methodology exists in the form of semiempirical Ca/S projections from a model developed by the Westinghouse Research and Development Center.[7] The model takes into account the chemistry and physics of the calcium-sulfur interactions in the FBC bed (viz., release of coal sulfur primarily as SO_2 and reaction with calcined sorbent to form $CaSO_4$). The model incorporates the following basic assumptions:[7]

- Release of sulfur from coal as SO_2 due to char and volatile combustion occurs uniformly throughout the combustor bed of AFBC units;

- The rate-limiting process for SO_2 capture in the bed is governed by diffusion within the sorbent particle itself; and

- Sorbent reactivity is a function of the bed calcining conditions and the degree of sulfation and is not independently affected by the residence time of sorbent particles in the bed.

The model also takes into account factors such as coal-ash alkali sulfur capture, the volume fraction of bed bubbles, bed voidage in the emulsion phase, the fraction of emulsion volume occupied by inerts, and the fraction of bed volume occupied by heat transfer surface. A complete description of the model is contained in Appendix C of Reference 7.

A summary table of Westinghouse model Ca/S projections as a function of SO_2 removal requirements and coal types is presented in Table 5.1-2. It should be noted that the specifications for the coal types in this table are the same as those used in the FBC-ITAR and this report. In addition, the Ca/S projections are based on an AFBC unit operating at 1550°F bed temperature, 4 feet bed depth, 6 feet/second superficial gas velocity, and 0.67 seconds residence time -- the same conditions as the ITAR "best system" design.

The Westinghouse projections are plotted in Figure 5.1-1 with the labels "optimistic" and "conservative" added to represent high reactivity/500 μm sorbent and average reactivity/1,000 μm sorbent, respectively. (For SO_2 removal efficiencies outside the range of Table 5.1-2, extrapolations were made using a power curve.) Sorbent reactivity is an intrinsic property of each stone and cannot, for practical purposes, be controlled. Low reactivity sorbents are not considered in this study because the high limestone feed rates and solid waste generation rates associated with their use make this option economically infeasible.

In-bed sorbent particle size is partly dependent on intrinsic stone properties such as feed particle size distribution and particle strength (i.e., resistance to attrition). In-bed particle size is also a function of solids residence time which in turn is determined by sorbent feed rate, bed volume, and recycle ratio. Thus the optimistic Ca/S projections identified above correspond to an FBC boiler feeding high reactivity limestone and operating with a longer solids residence time and/or a low-strength stone.

TABLE 5.1-2. WESTINGHOUSE PROJECTIONS FOR REQUIRED Ca/S RATIOS[a]

Sorbent Reactivity Category	High		Medium	
Average Bed Particle Diameter (Surface Mean), μm	500	1000	500	1000
SO_2 Emission Control Standard:				
(Percent Sulfur Removal)				
	Bituminous High-Sulfur Coal (3.5 wt. Percent S)			
Stringent (90)	2.8	3.5	3.4	4.3
Intermediate (85)	2.5	2.9	2.9	3.7
Moderate (78.7)	2.1	2.5	2.5	3.1
	Bituminous Low-Sulfur Coal (0.9 wt. Percent S)			
Stringent & Intermediate (84.7)	2.4	2.8	2.9	3.6
Moderate (75)	1.9	2.3	2.3	2.9
	Western Subbituminous Coal (0.6 wt. Percent S)			
Stringent & Intermediate (84.0)	1.1	1.3	1.3	1.7
Moderate (75)	0.7	0.9	1.0	1.2

[a] Source: Reference 7.

The conservative Ca/S projections correspond to average reactivity limestone, a shorter residence time, and/or high-strength stone. Since these conditions effectively cover the range of expected FBC boiler conditions, the actual rates for a given site should fall somewhere in between.

The data and information shown in Figure 5.1-2 demonstrate that the optimistic and conservative Westinghouse projections for Ca/S (as a function of SO_2 removal) form an envelope which contains most of the individual performance data points for industrial-scale AFBC units of conventional bed design. This agreement lends support to the use of the Westinghouse model Ca/S projections to estimate limestone requirements for model FBC boilers.

It should be noted that the outstanding SO_2 removal performance of the TVA 20 MW_e pilot plant operating with solids recycle may be aided by the higher freeboard of this unit. Freeboard height at the TVA unit is over 20 feet compared to near 10 feet for a typical industrial fluidized bed boiler. The higher freeboard allows more time for SO_2 capture by entrained sorbent, effectively increasing the in-bed gas residence time. Adjustment for this difference would tend to bring the TVA data within the Westinghouse envelope and closer to the optimistic projection. However, at this time, the impact of freeboard height on SO_2 removal is not defined well enough to make a quantitative adjustment.

The high Ca/S ratios observed in the Georgetown University tests may be explained in part by the low sorbent reactivity. More likely, these high Ca/S ratios reflect the design flaws and operational practices (e.g., the fluidized bed level was controlled by limestone addition) of a first-generation unit.[9] This unit is included for comparison, however, because it is one of the few commercial industrial FBC systems for which data are available.

In view of the fact that the Westinghouse model for Ca/S projections is a rigorous model which (1) adequately accounts for sulfur capture by coal-ash alkali species and (2) is in reasonable agreement with performance data from large operating systems, it is entirely appropriate to utilize the model results for purposes of cost algorithm development. The Westinghouse model is the best instrument currently available for projecting required Ca/S ratios as a function of SO_2 removal efficiency over the studied range of coal types and industrial FBC boiler operating conditions.

5.2 ALGORITHM DEVELOPMENT

The cost data in the FBC ITAR were based on a combination of FBC boiler vendor cost estimates, estimates developed by GCA for the limestone and spent solids handling and storage areas (based on vendor-supplied cost data), and guidelines developed by PEDCo for conventional boilers.[8] These data were used to develop capital and operating cost estimates for industrial AFBC boilers ranging in size from 30 to 200 million Btu/hr and feeding coals ranging from low sulfur western subbituminous to high sulfur eastern bituminous. It should be noted that Westinghouse has also developed cost estimates for FBC boilers, based in part on their Ca/S projection model. However, the cost sources for these estimates are Westinghouse in-house cost files (for the boiler and solids handling equipment) and literature references. The ITAR cost estimates are considered superior for the purposes of this study because (1) the boiler cost estimates were provided directly by commercial FBC vendors, and (2) data in the Westinghouse in-house cost files are not easily verified or referenceable. However, combining the ITAR cost data base with the Westinghouse model Ca/S projections takes advantage of the strengths of both data sets and provide the best basis currently available for developing FBC cost algorithms.

Details of the development history and modifications to the FBC cost algorithms are contained in Appendix A. The final form of the algorithm, as used in this report, is presented in Table A-1. Algorithm terms and units are explained in Table A-2.

The battery limits of the plant for which the algorithm applies are from, but not including, the coal receiving equipment and to, and including, the stack and onsite spent solids storage (on a temporary basis) equipment. It is assumed that spent solids are hauled by truck to an offsite landfill; the cost of this haulage is reflected in the solid waste disposal fee. A boiler feedwater treatment facility is included in the costs but steam piping to and from the process area is not. Battery limits include a primary cyclone for solids recycle but not a final particulate control device. No provisions are included for control of NO_x emissions below those levels characteristic of conventional AFBC technology.

The algorithm applies to coals ranging from high sulfur eastern bituminous to low sulfur western subbituminous (lignites are not included). Other applicable limits are:

- Boiler size: 30 - 400 million 10^6 Btu/hr heat input capacity
- Coal sulfur content: 0.6 - 3.5 wt. percent, as received basis
- Coal heating value: 9,600 - 13,800 Btu/lb, as received basis
- Coal ash content: 5.40 - 10.58 wt. percent, as received basis
- Coal moisture content: 2.83 - 20.8 wt. percent
- SO_2 removal efficiency: 56 - 90 percent
- Ca/S ratio: 0.8 - 4.2

Extrapolations outside these ranges should be made with caution; the results will have greater uncertainty than results within the indicated limits. It should be noted that these ranges apply only to the developed FBC cost algorithm. Although they represent typical conditions for industrial FBC boiler applications, they in no way stand for limitations to those applications.

5.3 COST COMPARISONS AMONG INDEPENDENT ESTIMATES

The performance data and results of Sections 3.0 and 4.0 indicate that the FBC cost algorithms and cost estimates of Section 6.0 are based on a realistic system design. To further test the validity of the FBC cost projections, it is desirable to compare them with independent estimates developed by other workers. In this section, the capital and annual cost estimates derived from the FBC algorithm are compared with independent estimates developed in the last few years by Combustion Engineering, Inc. (CE)[12], Foster Wheeler Development Corporation (FW)[13], Westinghouse Research and Development Center (W)[14], and Pope, Evans and Robbins, Inc. (PER)[15], as reported in literature sources. In addition, capital and operating costs for an installed and operating coal-fired FBC unit were provided by Johnston Boiler Company (JB).[16] With the exception of W, these companies currently offer commercial industrial-size FBC boilers.

Most of the vendor estimates identified above were developed for large capacity (greater than 200 million Btu/hr) boilers operating on high sulfur eastern coal in an industrial setting. In most instances, SO_2 emissions are controlled to a level of approximately 1.2 $lb/10^6$ Btu and PM emissions are controlled to near 0.05 $lb/10^6$ Btu. This set of conditions corresponds closely to the FBC boiler design case of 80 percent SO_2 removal on a Type H coal, as identified in Table 6.2-2. The exceptions to this rule are the JB costs which represent a 50 million Btu/hr boiler controlling SO_2 emissions to a 2.6 lb/million Btu limit.

The capital and operating costs developed by CE, FW, W, PER, and JB have been adjusted to achieve a consistent basis with the FBC algorithm projections so that valid comparisons can be made. The details of these adjustments have been summarized in Appendix C. After adjustments, the resulting capital and annual costs have been normalized on the basis of heat input capacity and plotted against boiler size in Figures 5.3-1 and 5.3-2, respectively. FBC algorithm costs corresponding to 80 percent SO_2 removal on a Type H coal have also been plotted on these figures for both optimistic and conservative Ca/S ratios. Error bands of ± 30 percent have been added to the algorithm capital and annual costs to represent the accuracy of the estimates (see Section 6.0).

For capital costs, Figure 5.3-1 demonstrates that the W, PER, and JB projections are well within the error limits of the FBC algorithm projections; the CE and FW estimates are near the limit of the upper error band. The actual algorithm projection for the JB case would be slightly lower than the band shown in the figure owing to the smaller limestone storage and spent solids handling equipment that correspond to a higher emission limit. The annual cost estimates plotted in Figure 5.3-2 show very good agreement among the FBC algorithm and the CE, FW, and W projections. No annual cost estimate could be developed for PER or JB because of a lack of information on O&M costs.

Overall, this comparison of five independent estimates with the FBC algorithm projections lends added validity to the algorithm as a cost estimating tool. Also, the fact that the independent estimates show some scatter with respect to the algorithm projections indicates that the algorithm is not biased either high or low.

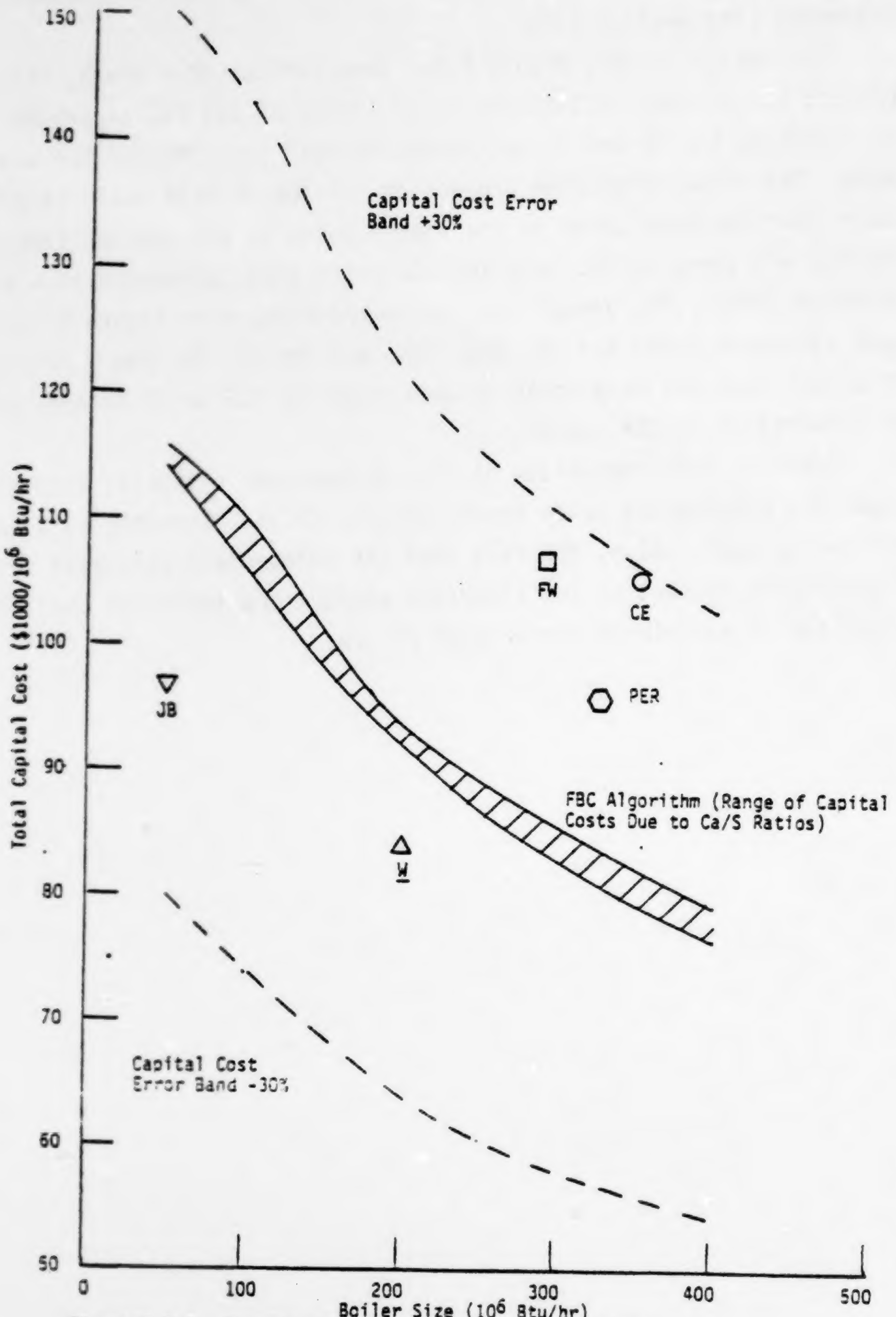

Figure 5.3-1. Comparison of Total Capital Cost Estimates

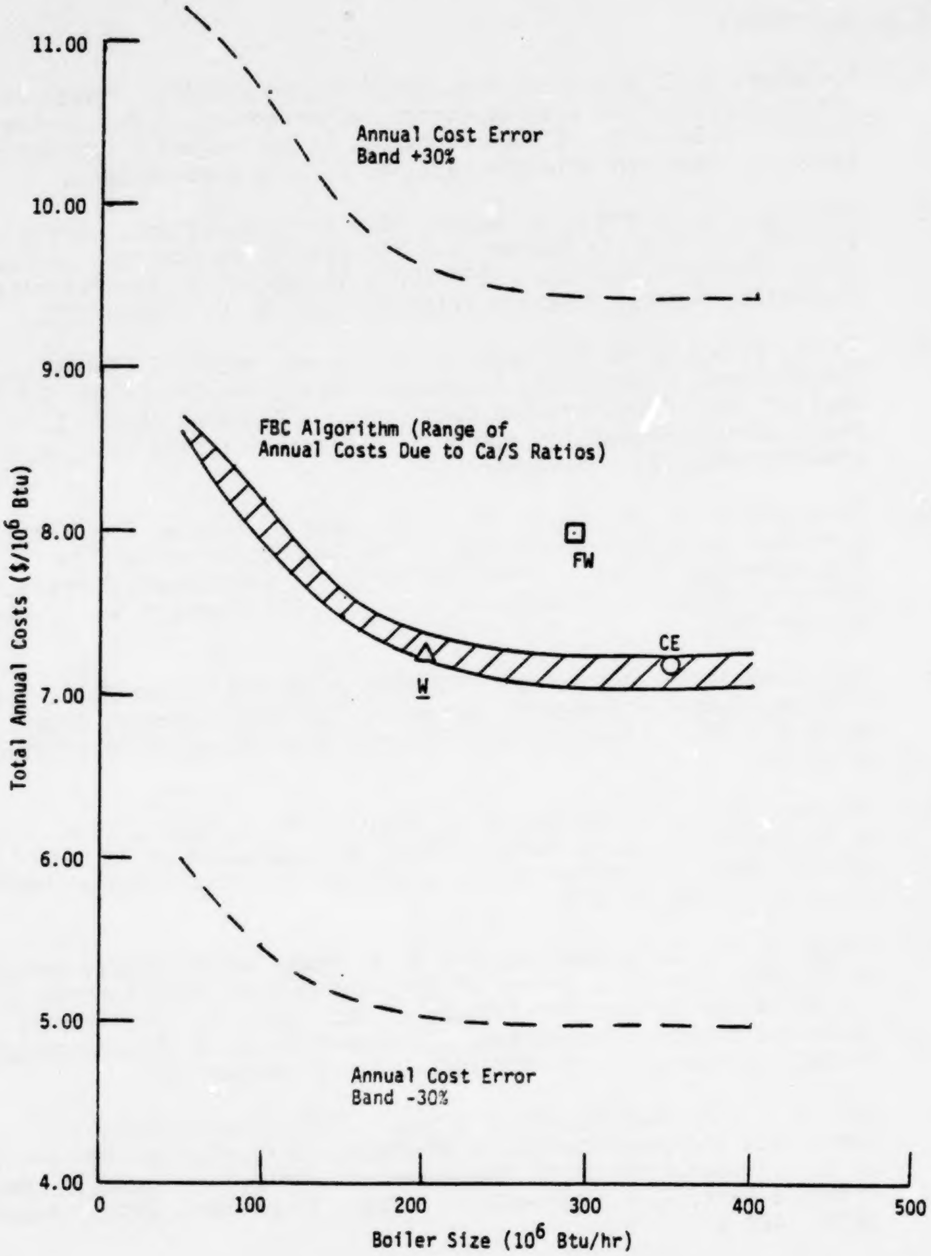

Figure 5.3-2. Comparison of Total Annual Cost Estimates

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6.0 ECONOMIC COMPETITIVENESS OF FBC TECHNOLOGY: IMPACT OF SO_2 EMISSION LIMITS

This section presents the capital and annual cost projections developed to assess the impact of alternative SO_2 emission standards on the relative competitiveness of industrial FBC steam generation systems. FBC costs are compared to two other SO_2 control alternatives: a conventional boiler equipped with an FGD system; and an uncontrolled conventional boiler burning low-sulfur compliance coal. The emphasis of this analysis is on trends and cost sensitivity. The costing techniques employed to develop the estimates presented in this section are consistent with budget-quality cost estimates (i.e., accurate to within ± 30 percent).

6.1 COSTING PREMISES

This report focuses on the cost competitiveness of industrial FBC technology as a function of SO_2 emission level stringency. Only coal-fired boilers have been assessed since SO_2 emission limits will have their greatest impact on FBC boilers operating on this fuel. While PM and NO_x emission limits are given due consideration, the objective of the analysis is to determine changes in relative cost competitiveness between these three SO_2 control alternatives as a function of SO_2 emission limits.

The SO_2 emission limits chosen for examination are 1.7, 1.2, and 0.8 lb $SO_2/10^6$ Btu. The 1.2 lb $SO_2/10^6$ Btu limit was chosen because it is currently the New Source Performance Standard (NSPS) for coal-fired boilers with heat input capacities greater than 250 million Btu/hr (40 CFR 60 Subpart D). The limits on either side of 1.2 were chosen to provide a reasonable range for the sensitivity analysis.

In order to meet these three SO_2 control levels on specified coals, FBC and conventional boiler/FGD options must achieve corresponding SO_2 removal efficiencies. The costs to achieve these efficiency levels, in conjunction with the emission limits identified above, will be used to assess the cost-competitiveness of FBC technology with FGD and low-sulfur coal options under various regulatory alternatives.

Allowable emissions of particulate matter (PM) and NO_x are maintained at consistent levels for all SO_2 control levels examined. PM and NO_x levels for both FBC and conventional coal-fired boilers are those levels recommended for new industrial steam generators under 40 CFR 60 Subpart D. These emission control levels and the methods for achieving control are summarized in Table 6.1-1.

6.1.1 Model Boilers

In this report, cost impacts are calculated using an analysis of the costs for model boilers and air pollution control systems. Model boilers and control system cost algorithms have been developed which represent typical industrial steam generating facilities for conventional systems.[1] The conventional system algorithms used in this study are presented in Reference 1; the algorithm for the FBC unit is described in Section 5 and Appendix A.

The model boiler sizes chosen for this study are 50, 100, 150, 250, and 400 million Btu/hr heat input; these capacities were chosen to provide a reasonable range of industrial boiler types and to include critical transition sizes with respect to PM and NO_x emissions. All of the conventional boilers are field-erected units, except the 50 million Btu/hr unit which is a shop-fabricated unit. FBC model boiler costs are based on a 30 million Btu/hr shop fabricated unit; a 75 million Btu/hr unit that was field erected from shop fabricated modules; and fully field erected 150 and 200 million Btu units. Costs for intermediate size units were interpolated using the cost algorithm. The 400 million Btu/hr facility consists of two 200 million Btu/hr boilers but a single train of limestone and spent solids storage and handing equipment. The conventional boiler types (viz., underfeed stoker, spreader stoker, and pulverized coal combustion) are specified in Table 6.1-1.

Explicit NO_x control methods are not required for FBC boilers to meet the emission limits identified in Table 6.1-1 because, as the data of

TABLE 6.1-1. NO_x AND PM EMISSION CONTROL LEVELS AND METHOD OF CONTROL

Boiler Size (10^6 Btu/hr)	Boiler Type	Emission Levels ($lb/10^6$ Btu)		Method of Control	
		NO_x	PM	NO_x	PM
50	Underfeed Stoker	0.6	0.05	Low excess air	Fabric Filter
50	AFBC	0.6	0.05	None	Fabric Filter
100	Spreader Stoker	0.6	0.05	Low excess air	Fabric filter
100	AFBC	0.6	0.05	None	Fabric filter
150	Spreader Stoker	0.6	0.05	Low excess air	Fabric filter
150	AFBC	0.6	0.05	None	Fabric filter
250	Pulverized Combustion	0.7	0.05	LEA/SCA[a]	Fabric filter
250	AFBC	0.7	0.05	None	Fabric filter
400	Pulverized Combustion	0.7	0.05	LEA/SCA[a]	Fabric filter
400	AFBC	0.7	0.05	None	Fabric filter

[a] LEA/SCA - low excess air in combination with staged combustion air.

Section 4 demonstrate, NO_x emissions from FBC units are consistenly below the 0.5 lb/10^6 Btu level specified for the smallest conventional boiler. A primary cyclone is included in the FBC boiler design but a final PM control device is necessary to reach the emission limits specified in the table.

6.1.2 SO_2 Control Alternatives

The SO_2 control alternatives selected for analysis in this report are: (1) an FBC boiler operating with limestone for SO_2 control (identified as FBC); (2) a conventional boiler equipped with a lime spray drying FGD system (identified as FGD); and (3) a conventional boiler firing low sulfur compliance coal (identified as CC).

It is assumed here that various SO_2 limitations identified above are based on continuous emission monitoring results. It is further assumed that the emission limits and removal requirements identified above are based on 30-day rolling averages. In order to comply with these requirements, compliance coal sulfur contents (on a lb $SO_2/10^6$ Btu basis) must be slightly lower than corresponding emission limits to allow for the natural variablity of coal sulfur content. A factor of 1.2 has been used in specifying the compliance coal corresponding to each emission limit (i.e., average SO_2 emissions are equal to the emission limit divided by 1.2). This factor is based on variability analyses of coal sulfur emissions data obtained from operating industrial boilers.[2] In most cases, a reference coal with the exact sulfur content required to meet the emission limit was not available; an available coal with a slightly lower sulfur content was specified (e.g., compliance coal with a sulfur content of 0.95 lb $SO_2/10^6$ Btu was specified to meet the 1.2 lb $SO_2/10^6$ Btu limit).

The SO_2 control alternatives, emission standards, and projected emission levels examined in this report are summarized in Table 6.1-2. For each FBC and FGD alternative in the table, two coal type options have been specified for comparison. The coal types used in this study are summarized in Table 6.1-3. Type H coal produces uncontrolled SO_2 emissions of 5.54 lb/10^6 Btu while Type F coal produces uncontrolled emissions of 2.85 lb/10^6 Btu. Of course the level of SO_2 removal efficiency required to meet a given

TABLE 6.1-2. SO_2 CONTROL ALTERNATIVES FOR MODEL BOILERS

Boiler Sizes (Million Btu/hr)	SO_2 Emission Limit ($lb/10^6$ Btu)	Control Alternative	SO_2 Control Technique	Coal[a] Type	% SO_2 Removal	Ca/S Ratio	SO_2 Emissions[b] ($lb/10^6$ Btu)
50, 100, 150, 250, 400	0.8	1A	FBC	H	90	4.30	0.55
		1B	FBC	H	90	2.80	0.55
		1C	FBC	F	80	3.20	0.57
		1D	FBC	F	80	2.20	0.57
		2A	FGD	H	90	1.68	0.55
		2B	FGD	F	80	1.29	0.57
		3	CC	A	-	-	0.60
50, 100, 150, 250, 400	1.2	1A	FBC	H	80	3.20	1.11
		1B	FBC	H	80	2.20	1.11
		1C	FBC	F	65	1.95	1.10
		1D	FBC	F	65	1.25	1.10
		2A	FGD	H	80	1.29	1.11
		2B	FGD	F	65	1.00	1.10
		3	CC	B	-	-	0.95
50, 100, 150	1.7	1A	FBC	H	75	2.75	1.39
		1B	FBC	H	75	1.85	1.39
		2	FGD	H	75	1.16	1.39
		3	CC	D	-	-	1.45

[a] Coal type specifications are summarized in Table 6.1-3

[b] SO_2 emissions are below the relevant emission limits to allow for the variability of coal sulfur content, FBC performance, and FGD performance. Compliance coal option emissions are slightly different than FBC and FGD option emissions due to reference coal sulfur specifications.

TABLE 6.1-3. COAL SPECIFICATIONS USED IN MODEL BOILER ANALYSIS[a]

Coal Type[d]	Fuel Price[b] ($/$10^6$ Btu)	Heating Value (Btu/lb)	Sulfur Content (Wt. %)	Sulfur Content (lb SO_2/ 10^6 Btu[c])	Ash Content (Wt. %)
Bituminous					
Type A	3.44	12,500	0.50	0.80	11.0
Type B	3.28	12,500	0.59	0.95	11.0
Type D	3.22	12,600	0.91	1.45	11.0
Type F	2.94	11,500	1.64	2.85	10.9
Type H	2.47	11,700	3.23	5.54	12.0
Subbituminous					
Type A	2.84	8,825	0.35	0.80	6.9
Type B	2.84	8,825	0.42	0.95	6.9

[a] Source: References 3, 4, and 5.

[b] 1990 levelized fuel prices in 1983 dollars.

[c] To obtain sulfur content in ng/J, multiply by 430.

[d] Coal specifications are based on average specifications for Midwest region.

emissions limit declines from Type H to Type F coal, as reflected in Table 6.1-2. These coal types are examined to illustrate the sensitivity of system costs to coal sulfur content and SO_2 removal efficiency requirements. For the FBC cases, two levels of Ca/S ratio are examined, corresponding to the optimistic and conservative Ca/S projections of Section 5.1.2, for each coal type. SO_2 removal efficiency levels for FBC and FGD alternatives were chosen to yield emission levels approximately equal to CC levels.

In the case of the 1.7 lb $SO_2/10^6$ Btu limit, boiler sizes of 250 million Btu/hr and above were not considered since the limit for this boiler category is already set at 1.2 lb $SO_2/10^6$ Btu (see 40 CFR 60 Subpart D).

The FGD system specified for this analysis is the lime spray drying system. This system was chosen over other FGD systems (e.g., dual alkali, lime/limestone, or sodium once-through wet scrubbing) because (1) the technology is being widely applied for SO_2 control among industrial boilers; (2) spray drying costs are representative of costs for other FGD technologies (e.g., once-through sodium and dual alkali FGD) throughout the studied size range; and (3) the technology is similar to FBC technology in its use of a calcium sorbent and production of a dry waste material.[1] Lime spray drying systems include a fabric filter as an integral part of their design and thus achieve combined PM and SO_2 control. Detailed specifications for this system, as well as other PM and NO_x control techniques are presented in Reference 1.

As mentioned above, lime spray drying costs are generally representative of FGD costs over the range of industrial boiler applications examined. For smaller boilers below about 200 million Btu/hr, sodium once-through wet scrubbing appears to be the low-cost alternative while for larger boilers above 300-350 million Btu/hr dual alkali wet scrubbing exhibits the lowest costs.[1] Throughout this range, dry lime scrubbing costs fall between the costs for these two wet scrubbing alternatives. In no case do the estimated annual costs for these three technologies differ by more than 15 percent. In view of this comparison, lime spray drying costs were chosen as most representative of industrial FGD costs in this boiler size range.

6.1.3 Coal Specifications

The largest operating and maintenance (O&M) cost for both conventional and FBC boilers is fuel. Table 6.1-3 presents the specifications and costs for the coals used in this analysis. The prices in this table are projections for 1990 delivered fuel prices expressed in January 1983 dollars.[3,4,5] These projections ignore the effects of inflation but assume that fuel prices will escalate in real terms. In addition, the fuel prices have been "levelized" over the life of the boiler (i.e., an equivalent constant price has been calculated after allowing for escalation and the time value of money). These fuel prices are used in this study to maintain consistency with other industrial model boiler cost analyses conducted within EPA.[1]

Direct O&M costs for the boilers and control devices are calculated using the algorithms referenced above. The key factors used in estimating annual O&M costs are the system capacity utilization, utility unit costs (steam, electricity, water), and unit costs for raw materials, waste disposal, and labor. In keeping with the above-mentioned model boiler cost analyses, non-fuel O&M costs are assumed to escalate at the same rate as inflation so that there is no increase in "real" costs. Capacity utilization is defined as the actual annual fuel consumption as a percentage of the potential annual fuel consumption at maximum firing rate. A value of 0.6 has been assumed in this study; this value corresponds to current practice as defined in other industrial boiler cost analyses.[1] Table 6.1-4 summarizes the utility and unit costs used in calculating annual O&M costs for the boilers and control equipment.

A complete description of the cost bases utilized for capital and annual cost calculations is presented in Appendix D.

6.2 COST COMPARISON RESULTS

Before discussing cost comparison results, it should be noted that the cost data on which both the FBC and conventional system cost algorithms are based come from respective ITAR cost estimates, which are considered

TABLE 6.1-4. UNIT COSTS USED IN MODEL BOILER CALCULATIONS[a]

Utilities	
Electricity	0.0503/kwh[b]
Water	0.0396/m^3 (\$0.15/10^3 gal)[c]
Steam	\$3.5/10^3 lb[d]
Raw Materials	
Na_2CO_3	\$0.169/kg (\$153/ton)[c,e]
Lime	\$0.098/kg (\$89/ton)[c,e]
Limestone	\$0.013/kg (\$8.5/ton)[c]
Labor	
Direct Labor	\$11.75/man-hour[f,g]
Supervision	\$15.28/man-hour[h]
Maintenance Labor	\$14.34/man-hour[i]
Waste Disposal	
Solids (Ash, Spray Dried Solids)	\$0.0198/kg (\$18/ton)[j]
Sludge	\$0.0198/kg (\$18/ton)[j]

[a] All costs in January 1983 $.

[b] Monthly Energy Review, April 1983.

[c] TVA, Technical Review of Dry FGD Systems and Economic Evaluation of Spray Dryer FGD Systems, February 1982.

[d] EPRI, Technical Assessment Guide, May 1982.

[e] Updated using ratio of commodity chemical price for January, 1983 to June, 1982 as given in the Chemical Marketing Reporter.

[f] Monthly Labor Review April, 1982.

[g] Average of wate rates for Chemical and Allied Products and Petroleum and Coal Products categories.

[h] Estimated at 30 percent over direct labor rate.

[i] Estimated at 22 percent over direct labor rate.

[j] Average of waste disposal rates from Economics of Ash at Coal Fired Power Plants, Oct. 1981, and EEA, Estimated Landfill Credit for Non-Fossil Fueled Boilers, October, 1980.

accurate to approximately ±30 percent. Thus the capital cost estimates in this report retain the same level of accuracy. In making comparisons between FBC and other technology options, however, the accuracy of capital cost differences may be better than ±30 percent. This is due to the fact that some equipment items are common to all algorithms and have been treated in the same manner (e.g., use of PEDCo data to estimate the cost of boiler feed pumps).

The accuracy of total annual cost estimates is also ± 30 percent. However, relatively little error is associated with comparisons of total O&M costs between technologies since (1) raw material and fuel requirements can be estimated with a high degree of accuracy (based on assumptions in most cases) and (2) the same unit costs have been used in estimating operating costs for each alternative (e.g., hourly labor rates, solid waste disposal rate, plant and payroll overhead). Therefore, annual cost error bands are primarily due to the error associated with annualized capital charges. On this basis, total annual cost comparisons between technology options are considered accurate to within about 15 percent over the boiler size range examined.

The accuracy limits for capital and operating costs should be borne in mind when reviewing the results discussion in this section and Sections 6.3 and 6.4. The absolute value of any single cost estimate is accurate only to within the error bands specified above.

6.2.1 Overall Results

Tables 6.2-1 to 6.2-3 summarize the annual cost estimates for the SO_2 control alternatives outlined in Section 6.1.2. The cost estimates have been grouped by SO_2 emission limitations so that alternatives can be compared with other alternatives of approximately equal SO_2 control stringency. The tables presented in Appendix B show how the boiler, NO_x control, SO_2 control, and PM control equipment costs contribute to overall capital and operating costs for each control alternative.

A review of the Appendix B cost summaries indicates that, for the FBC options, Ca/S ratios can vary by as much as 50 percent for each option due

TABLE 6.2-1. TOTAL ANNUAL COSTS FOR SO_2 CONTROL OPTIONS AT 1.7 LB/10^6 BTU EMISSION LIMIT
TOTAL ANNUAL COSTS ($1000)[a]

Boiler Size (Million Btu/hr)	Fluidized Bed Combustion[a,c] 75%/Type H[d]	Conventional Boiler/FGD[c] 75%/Type H	Conventional Boiler/ Compliance Coal Type D
50	2,278	2,282	2,076
100	4,228	4,019	3,931
150	5,961	5,554	5,562

[a] January 1983 dollars.

[b] Based on conservative Ca/S ratios (see Appendix B).

[c] Only Type H coals are examined for these options since firing a Type F coal would correspond to only 50 percent SO_2 removal, a level which is not encountered in typical industrial boiler applications.

[d] SO_2 removal percentage/coal type.

TABLE 6.2-2. TOTAL ANNUAL COSTS FOR SO_2 CONTROL OPTIONS AT 1.2 LB/10^6 BTU EMISSION LIMIT TOTAL ANNUAL COSTS ($1000)[a]

Boiler Size (Million Btu/hr)	Fluidized Bed Combustion[b]		Conventional Boiler/FGD		Conventional Boiler/ Compliance Coal	
	80%/Type H[c]	65%/Type F	80%/Type H	65%/Type F	Type B Sub	Type B Bit
50	2,297	2,326	2,301	2,330	2,266	2,160
100	4,291	4,316	4,053	4,124	3,915	4,004
150	6,024	6,056	5,604	5,727	5,519	5,667
250	9,510	9,586	9,504	9,723	9,332	9,709
400	15,293	15,451	13,810	14,183	13,656	14,342

[a] January 1983 dollars.

[b] Based on conservative Ca/S ratios (see Appendix B)

[c] SO_2 removal percentage/coal type.

TABLE 6.2-3. TOTAL ANNUAL COSTS FOR SO_2 CONTROL OPTIONS AT 0.8 LB/10^6 BTU EMISSION LIMIT TOTAL ANNUAL COSTS ($1000)[a]

Boiler Size (Million Btu/hr)	Fluidized Bed Combustion[b]		Conventional Boiler/FGD		Conventional Boiler/ Compliance Coal	
	90%/Type H[c]	80%/Type F	90%/Type H	80%/Type F	Type A Sub	Type A Bit
50	2,341	2,355	2,355	2,358	2,266	2,140
100	4,393	4,370	4,154	4,173	3,915	4,088
150	6,177	6,159	5,751	5,797	5,519	5,793
250	9,765	9,753	9,743	9,834	9,332	9,922
400	15,702	15,695	14,183	14,354	13,656	14,682

[a] January 1983 dollars.

[b] Based on conservative Ca/S ratios (see Appendix B)

[c] SO_2 removal percentage/coal type.

to differences between the optimistic and conservative projections, as explained in Section 5.1-2. Despite this large difference in Ca/S ratios, annual costs differ by only 1 to 4 percent over the range of boiler sizes and SO_2 emission limits examined. This is due to the fact that limestone raw material costs and solid waste disposal costs are a relatively small fraction of overall annual costs. Thus Ca/S ratios have only a small impact on total annual FBC system costs. In light of this small difference, and the desire to develop conservative estimates of FBC technology costs (i.e., to err on the high side), only the conservative Ca/S ratios results will be considered in the discussion of this and following sections of the report.

A careful examination of the cost estimates summarized in Tables 6.2-1 to 6.2-3 reveals several important overall results:

- For the SO_2 control options meeting a 1.2 lb/10^6 Btu limit, the annual costs for both the FBC and FGD alternatives are lower (2 to 3 percent) for the Type H coal options than the Type F coal options. This is because the added fuel charges for the lower sulfur content, but more expensive, Type F coal outweigh the capital and operating cost savings which result from lower limestone feed and solid waste disposal requirements.

- For the 0.8 lb $SO_2/10^6$ Btu cases, this same trend applies for the FGD alternatives but is reversed for the FBC alternatives above 50 million Btu/hr heat input. Due to the higher Ca/S ratios associated with 90 percent SO_2 removal in an FBC unit, a crossover point is reached between 50 and 100 million Btu/hr heat input at which lower overall annual costs are incurred by removing only 30 percent of the SO_2 from a Type F coal. This crossover point is not observed for the FGD alternatives in the studied ranges because of the lower Ca/S ratios associated with this technology.

- When comparing bituminous to subbituminous Type A and B coals, lower annual costs are incurred in most cases by firing the subbituminous coals since their lower fuel costs more than offset the higher boiler capital costs due to lower heating values. The exceptions to this rule are the 50 million Btu/lb boilers where low fuel use rates do not generate sufficient fuel cost savings to offset higher capital costs. For small boilers meeting 1.2 and 0.8 lb $SO_2/10^6$ Btu emission limits firing bituminous coal results in lower overall annual costs. This advantage disappears at the 100 million Btu/hr size and above.

- When comparing the low annual cost options for FBC with the low annual cost options for FGD and CC, the FBC technology costs are shown to be comparable to the costs for the other alternatives over the boiler size range and SO_2 emission range examined. That is, annual cost differences between options do not exceed 15 percent, which is within the overall accuracy of the annual cost comparisons.

- Capital costs for the three SO_2 control options are also comparable (i.e., within ±30 percent) for boilers above 50 million Btu/hr heat input. For small boilers near 50 million Btu/hr, CC capital costs are significantly lower than those for FBC units.

6.2.2 FBC Competitiveness Across SO_2 Emission Limits

In order to gain perspective on the influence of SO_2 emission limits on relative economic competitiveness, FBC annual costs are compared with costs for FGD and compliance coal in Table 6.2-4. Negative values in this table represent cases where FBC is projected to be more attractive than the other options. Total annual costs for these technology options are also plotted in Figures 6.2-1 and 6.2-2 as a function of SO_2 emission rates (equivalent to coal sulfur contents for compliance coals). The focus of this analysis is on annual costs since both plant owners and various boiler/fuel choice analysis models make their selection among SO_2 control alternatives primarily on the basis of total annual costs.

TABLE 6.2-4. FBC ANNUAL COST COMPETITIVENESS WITH FGD AND COMPLIANCE COAL AS A FUNCTION OF EMISSIONS LIMIT

	Boiler Size (Million Btu/hr)	SO_2 Emission Limit ($lb/10^6$ Btu)		
		1.7	1.2	0.8
FBC vs FGD[c]	50	-0.2	-0.2	-0.6
	100	5.2	5.9	5.8
	150	7.3	7.5	7.4
	250	-	0.1	0.2
	400	-	10.7	10.7

	Boiler Size (Million Btu/hr)	SO_2 Emission Limit ($lb/10^6$ Btu)		
		1.7	1.2	0.8
FBC vs.[c] Compliance Coal	50	9.7[b]	6.3	9.4
	100	7.6	9.6	12.2
	150	7.2	9.2	11.9
	250	-	1.9	4.6
	400	-	12.0	15.0

[a] Values correspond to (FBC annual costs/FGD annual costs) x 100 - 100.

[b] Values correspond to (FBC annual costs/compliance coal costs) x 100 - 100.

[c] Annual cost for each alternative corresponds to lowest annual cost option in Appendix B tables; FBC costs are based on conservative Ca/S ratios.

FIGURE 6.2-1

FBC ANNUAL COST COMPETITIVENESS WITH FGD

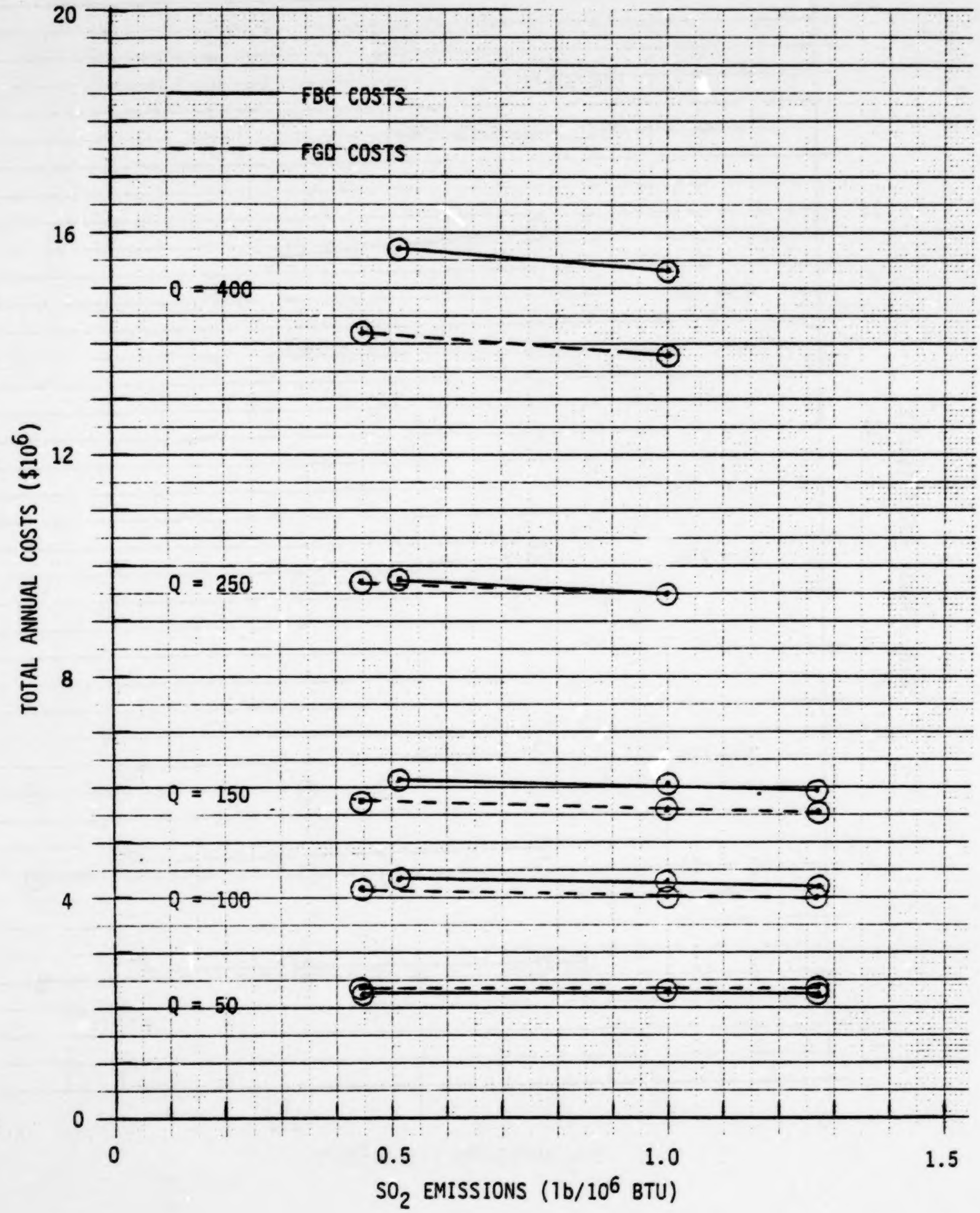
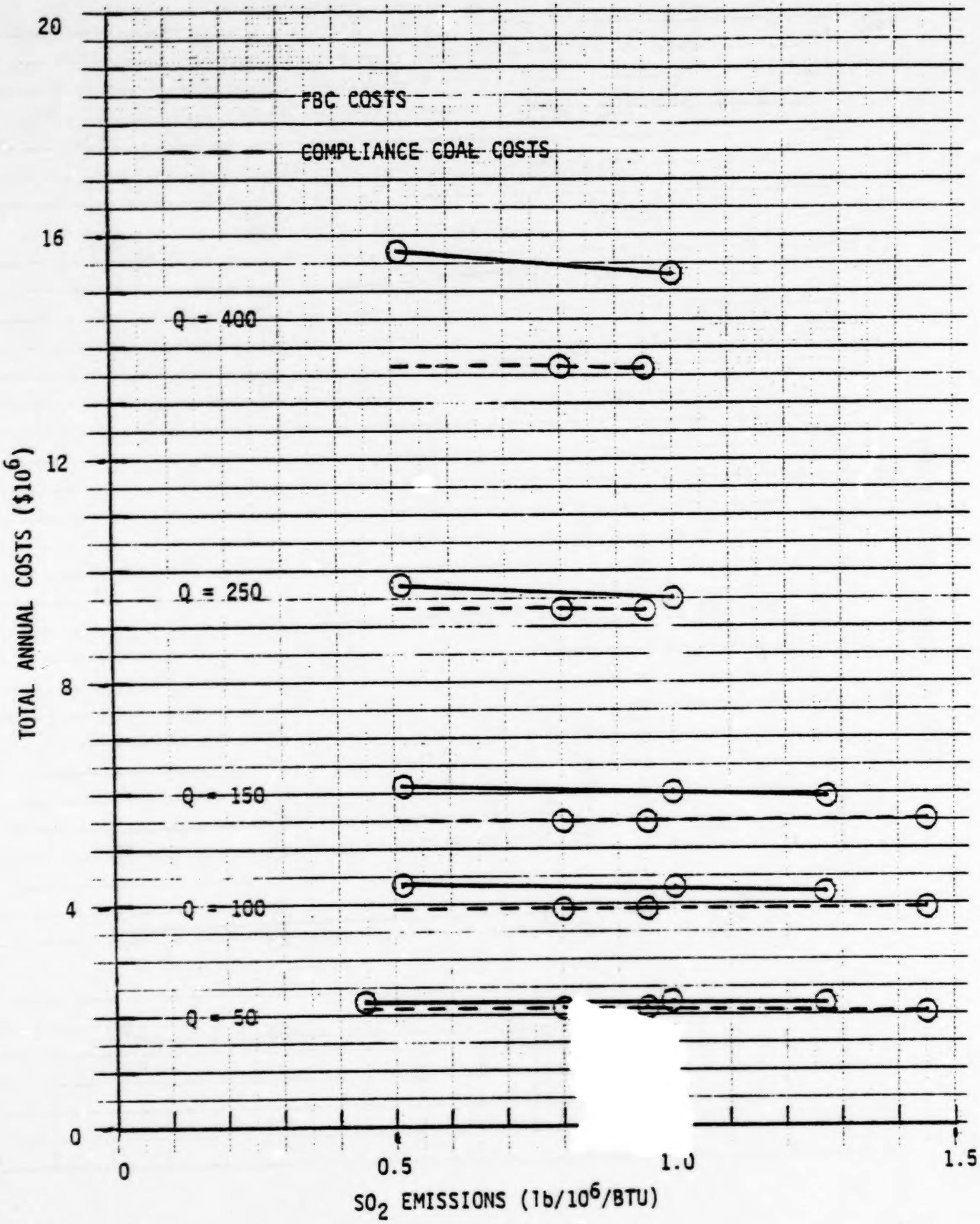


FIGURE 6.2-2

FBC ANNUAL COST COMPETITIVENESS WITH COMPLIANCE COAL

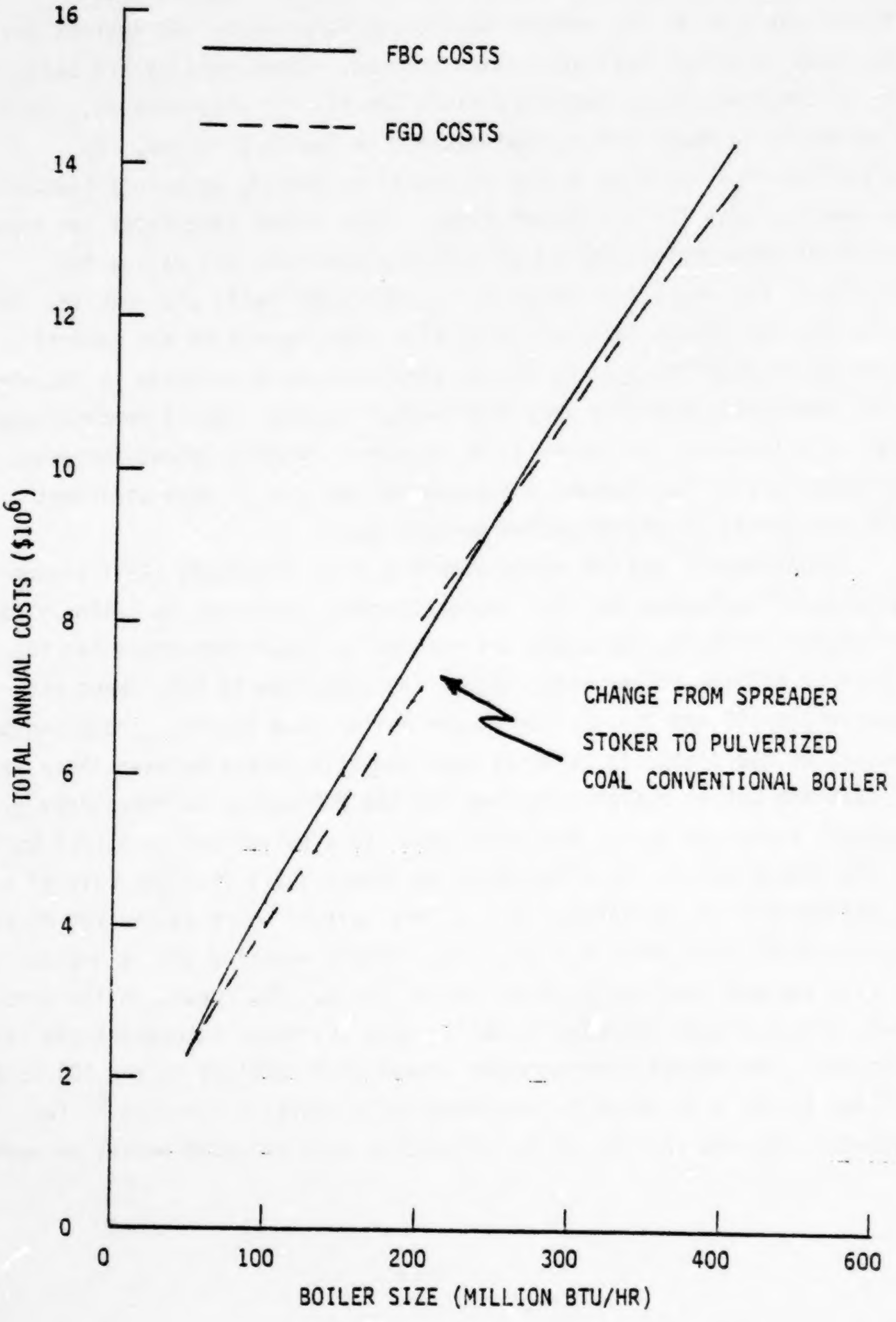


The information in Table 6.2-4 and Figure 6.2-1 indicates that FBC competitiveness relative to FGD remains nearly constant as the SO_2 emissions limitation becomes stricter for all boiler sizes. Thus FBC cost effectiveness as an SO_2 control technology relative to FGD systems does not change as emission level stringency changes. These results are based on the use of conservative or high Ca/S ratio for the FBC alternatives. It is interesting to note that for optimistic, or low Ca/S ratios, FBC competitiveness relative to FGD increases as the SO_2 emissions limitations becomes stricter for all boiler sizes. Thus larger incentives for research and development efforts aimed at lowering required Ca/S ratios for industrial FBC units will occur as SO_2 emission limits are reduced. This trend for optimistic Ca/S ratios is also consistent with the general observation that FBC systems can be very attractive relative to FGD when plant operators have only very high sulfur (greater than 4 percent) coal available for use. In general, FBC economic competitiveness increases as the mass rate of SO_2 removal increases, either due to more stringent emission limits or higher sulfur content coal.

Comparing FBC and FGD costs within a given emissions limit category, Table 6.2-4 indicates that FBC competitiveness increases as boiler size decreases. In fact, FBC costs are marginally lower than those for FGD units at the 50 million Btu/hr size range. The exception to this trend occurs between the 150 and 250 million Btu/hr boiler size levels. The principal reason for the change in relative cost competitiveness between these levels is that the boiler design specified for the FGD option switches from a spreader stoker boiler at the lower level to a pulverized coal (PC) boiler at the higher level. As illustrated in Figure 6.2-3 (for the case of a 1.2 lb/million Btu SO_2 emissions limit), this switch occurs at the 200 million Btu/hr boiler size level for the model boilers examined and is accompanied by a 13 percent increase in total annual costs. FBC costs, on the other hand, show a steady increase as boiler size increases throughout the range examined. The change from spreader stoker to PC boilers in the 200 to 300 million Btu/hr size range is consistent with industry practice.[6] Two secondary reasons for the shift in relative cost competitiveness between the

FIGURE 6.2-3

FBC AND FGD ANNUAL COSTS FOR A 1.2 LB $SO_2/10^6$ BTU EMISSION LIMIT



150 and 250 million Btu/hr boiler size levels are: (1) the cost of NO_x emission controls on the conventional boiler changes from a negative cost (due to effect of LEA use on stoker boiler fuel savings) to a net positive cost associated with the use of LEA/SCA on PC boilers; and (2) multiple boilers are specified for the FBC option above the 200 million Btu/hr range which results in a slight decrease in annual costs (less than 1 percent).

Figure 6.2-3 also shows that FGD option annual costs generally increase at a slower rate than FBC option costs as boiler size increases. As a result, FBC cost competitiveness decreases as boiler size increases, except in the case noted above.

Assessment of the information in Table 6.2-4 and Figure 6.2-2 concerning FBC cost competitiveness relative to compliance coal combustion indicates that most of the same trends apply: (1) relative cost competitiveness between the two alternatives remains nearly constant over the studied range of SO_2 emission limits and (2) FBC cost competitiveness decreases slightly as boiler size increases except in the range of 150 to 250 million Btu/hr. This latter behavior is illustrated in Figure 6.2-4. As discussed earlier, the principal reason for the change in relative cost competitiveness between these levels is the switch from spreader stoker to PC boilers for the compliance coal option.

There is a slight decrease in FBC cost competitiveness relative to CC as the emission limit is reduced from 1.2 to 0.8 $lb/10^6$ Btu. This is due primarily to the fact that FBC annual costs increase with decreasing emission levels (owing to higher capital and operating costs for limestone and spent solids disposal) while compliance coal prices either do not change between Type A and B coals (for subbituminous coals) or change only slightly (for bituminous coals). An expanded discussion of the impact of coal prices on FBC competitiveness is presented in Section 6.3.

Unlike the FBC-FGD cost comparison, FBC competitiveness relative to CC remains constant as the SO_2 emission limit decreases if the optimistic Ca/S ratios are used. The only case for which FBC costs appear marginally lower than CC costs at the lower Ca/S ratios occurs at the 250 million Btu/hr boiler level.

FIGURE 6.2-4
FBC AND COMPLIANCE COAL ANNUAL COSTS FOR A
1.2 LB $SO_2/10^6$ BTU EMISSION LIMIT

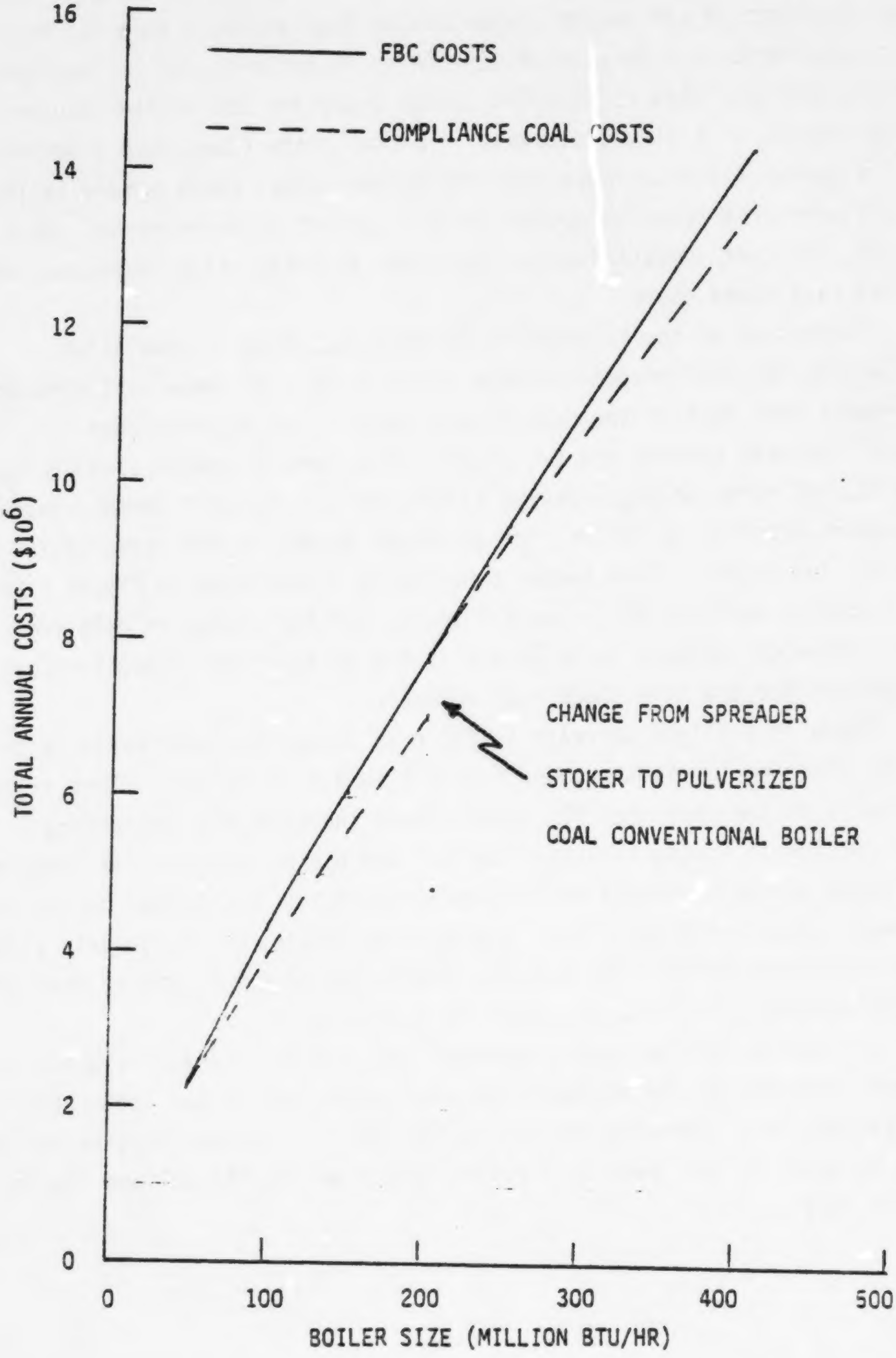


Table 6.2-5 provides an overview of the capital cost competitiveness of FBC with the FGD and CC alternatives. It shows that capital cost competitiveness remains relatively constant among alternatives as the emission limit varies. FBC capital costs are most attractive at the larger boiler sizes. FBC capital costs are significantly above those of CC alternatives at the 50 million Btu/hr level.

6.2.3 FBC Competitiveness Based on SO_2 Percent Removal Requirements

A second type of SO_2 emission limitation which currently applies to electric utility boilers above 250 million Btu/hr heat input capacity [Subpart Da (40 CFR Part 60)] is a requirement for a specific level of SO_2 removal efficiency. To evaluate this type of limitation, FBC annual costs are compared with FGD costs for equal SO_2 removal performance levels in Table 6.2-6. Not surprisingly, the data follow the same trends identified earlier for an emissions limit measured in lb $SO_2/10^6$ Btu heat input. FBC competitiveness vis-a-vis FGD remains relatively unchanged over the studied range of SO_2 percentage removal requirements. If the optimistic Ca/S ratios are used for the FBC alternatives, FBC competitiveness increases as SO_2 removal levels become more stringent.

As was the case in Table 6.2-4, FBC competitiveness in Table 6.2-5 relative to FGD increases as boiler size decreases, all other things being equal. The same factors as cited above also account for the change in relative competitiveness between the 150 and 250 million Btu/hr boiler size categories.

The capital cost figures shown in Table 6.2-7 indicate that FBC competitiveness relative to FGD on a capital cost basis remains constant as SO_2 removal efficiency varies. FBC capital costs are slightly below those of the FGD alternatives for 250 and 400 million Btu/hr boilers.

6.3 CONDITIONS UNDER WHICH FBC IS ECONOMICALLY FAVORED

One of the objectives of this study is to identify those conditions under which FBC is economically favored over a conventional boiler/FGD

TABLE 6.2-5. FBC CAPITAL COST COMPETITIVENESS WITH FGD AND COMPLIANCE COAL AS A FUNCTION OF EMISSIONS LIMIT

	Boiler Size (Million Btu/hr)	SO_2 Emission Limit (lb/10^6 Btu) 1.7	1.2	0.8
	50	12.7[a]	12.8	12.8
	100	8.1	9.9	4.0
FBC vs FGD[c]	150	9.5	10.1	4.5
	250	-	-2.3	-7.2
	400	-	1.1	-4.5
	Boiler Size (Million Btu/hr)	SO_2 Emission Limit (lb/10^6 Btu) 1.7	1.2	0.8
	50	39.3[b]	30.8	41.5
	100	21.1	10.9	6.3
FBC vs.[c] Compliance Coal	150	17.9	9.1	4.7
	250	-	-0.1	-4.1
	400	-	3.3	-1.4

[a] Values correspond to (FBC capital costs/FGD capital costs) x 100 - 100.

[b] Values correspond to (FBC capital costs/compliance coal capital costs) x 100 - 100.

[c] Capital cost for each alternative corresponds to lowest annual cost option in Appendix B tables; FBC costs are based on conservative Ca/S ratios.

TABLE 6.2-6. FBC ANNUAL COST COMPETITIVENESS WITH FGD AS A FUNCTION OF SO_2 PERCENT REMOVAL REQUIREMENT[b]

Boiler Size (Million Btu/hr)	SO_2 Removal Efficiency (Percent) 65	 75	 80	 90
50	-0.2[a]	-0.2	-0.2	-0.6
100	4.7	5.2	5.9	5.8
150	5.7	7.3	7.5	7.4
250	-1.4	-	0.1	0.2
400	8.9	-	10.7	10.7

[a] Values correspond to [(FBC annual costs/FGD annual cost) x 100 - 100].

[b] Annual cost for each alternative corresponds to lowest annual cost option in Appendix B tables; FBC costs are based on conservative Ca/S ratios.

TABLE 6.2-7. FBC CAPITAL COST COMPETITIVENESS WITH FGD AS A FUNCTION OF SO_2 PERCENT REMOVAL REQUIREMENT[b]

Boiler Size (Million Btu/hr)	SO_2 Removal Efficiency (Percent) 65	75	80	90
50	12.6[a]	12.7	12.8	12.8
100	6.9	8.1	9.9	4.0
150	6.1	9.5	10.1	4.5
250	-6.4	-	-2.3	-7.2
400	-4.9	-	-1.1	-4.5

[a] Values correspond to [(FBC capital costs/FGD capital cost) x 100 - 100].

[b] Capital cost for each alternative corresponds to lowest annual cost option in Appendix B tables; FBC costs are based on conservative Ca/S ratios.

system or compliance coal. The cost information in Tables 6.2-1 through 6.2-3 indicate that FBC is economically equivalent on an annual cost basis to FGD and compliance coal combustion for the cases under consideration in view of the overall accuracy of the annual cost comparisons (i.e., ± 15 percent).

To be significantly favored over the other alternatives, FBC should be approximately 15 percent less expensive on an annual cost basis. This assumes that there is a high probability that the true cost differential between two technologies will be within 15 percent of the cost differential estimated by the algorithms. Using this criterion of a 15 percent cost differential, key parameters can be varied in the annual cost basis to identify those conditions under which FBC is a clear favorite.

A 150 million Btu/hr boiler and 0.8 lb SO_2/MM Btu emission limit have been chosen as the basis of this analysis. The cost data of the previous sections show that FBC is least competitive, in most cases, at the 150 million Btu/hr boiler size. Thus the parameter adjustments required for the 150 million Btu/hr boiler will be generally greater than those required for other boiler sizes. The 0.8 lb $SO_2/10^6$ Btu standard has been chosen because it is the most stringent control limit considered in this study as regards both final emissions and percent reductions as well as the annual cost savings required.

6.3.1 FBC Versus FGD

As indicated in Table 6.2-3, in order to be 15 percent less expensive than FGD, the FBC option annual costs should be no more than $4,888,000 (i.e., (1.00-0.15) x $5,751,000). The annual costs for the FBC option in this case are summarized in Table 6.3-1. To achieve the target annual cost identified above, a cost savings of $1,271,000 is required. A study of Table 6.3-1 shows that FBC limestone and solid waste disposal costs could drop to zero, simultaneously, and only reach about one-fifth of the desired annual cost savings. This is not possible, of course, since the minimum theoretical Ca/S molar ratio for SO_2 capture is 1.0. The point here is that

TABLE 6.3-1. DETAILED ANNUAL COST BREAKDOWN FOR FBC (BASIS: 150x10^6 BTU/HR, TYPE F coal, 80 PERCENT SO_2 REMOVAL, Ca/S = 3.20, JAN 1983 $)

	FBC Boiler	Baghouse	Total
Direct Operating Cost			
Direct Labor	$ 217,000	$ 19,000	$ 236,000
Supervision	92,000	-	92,000
Maintenance Labor	86,000	13,000	99,000
Replacement Parts	213,000	12,000	225,000
Electricity	230,000	39,000	269,000
Process Water	19,000	-	19,000
Fuel	2,319,000	-	2,319,000
Limestone	53,000	-	53,000
Waste Disposal	142,000	15,000	157,000
Total Direct Cost	3,371,000	98,000	3,470,000
Overhead			
Payroll	65,000	6,000	71,000
Plant	158,000	11,000	169,000
Total Overhead Cost	223,000	17,000	240,000
Capital Charges			
Capital Recovery	1,677,000	164,000	1,841,000
Working Capital Interest	46,000	2,000	48,000
Miscellaneous	510,000	50,000	560,000
Total Capital Charges	2,233,000	216,000	2,449,000
Total Annual Costs	$ 5,827,000	$ 332,000	$ 6,159,000

reducing the Ca/S ratio alone will not have a significant impact on FBC competitiveness relative to FGD.

The two largest factors influencing annual FBC costs are fuel charges and capital costs. Since the FBC and FGD alternatives use the same fuel at the same rate (i.e., boiler efficiencies for FBC and conventional boilers are assumed equivalent), a comparative cost savings based on fuel charges is not possible. With respect to capital costs, the information in Appendix D indicates that model boiler turnkey costs are multiplied by a factor of 0.1715 to calculate the annual costs due to capital recovery and miscellaneous costs. Thus a turnkey cost reduction of \$7.41 million (\$1,271,000 ÷ 0.1715), or 51 percent would be required to lower total FBC annual costs to a level 15 percent below FGD costs. Conversely, FGD capital costs would have to rise by 73 percent to accomplish the same effect. Neither of these changes, at least of this magnitude, are likely to occur in the foreseeable future as a result of technological developments.

6.3.2 FBC Versus Compliance Coal

Annual FBC cost reductions relative to compliance coal combustion must be even greater than those relative to FGD. To achieve the same 15 percent annual cost advantage over the CC option at the base conditions, FBC costs should be \$4,691,000 per year, or a reduction of \$1,468,000.

Table 6.3-1 results indicate that either fuel charges or capital costs, or both, should be reduced to effect this cost reduction. In the case of fuel charges, a differential of $\$1.86/10^6$ Btu would be sufficient to make FBC a clear economic favorite over compliance coal. This differential could be achieved either by lowering the unit cost of the Type H coal burned in the FBC unit or raising the unit cost of the Type A coal burned in the conventional spreader stoker boiler, or a combination thereof. This corresponds to a 63 percent reduction of unit coal costs for the FBC option or a 65 percent increase for the compliance coal unit cost.

As with the FGD comparison, the relative turnkey capital costs for the FBC and CC options could be shifted to achieve the targeted FBC annual cost advantage. This target translates to a \$8.56 million turnkey capital cost differential which corresponds to a 59 percent reduction of FBC costs or a 62 percent increase for CC costs, or a combination of the two. Again, as with the earlier discussion concerning FGD costs, the likelihood of cost changes of this magnitude occurring in the foreseeable future as a result of coal market or technological changes is quite remote.

The figures presented in this section are not projections or predictions of changes that will occur among the three technology alternatives. Rather, the calculations are meant to illustrate the length to which unit costs and turnkey capital costs would have to change to make the FBC option a clear-cut favorite over FGD and CC for a 150 million Btu/hr boiler operating to meet a 0.8 lb $SO_2/10^6$ Btu limit on a continuous basis. Relative changes of a similar magnitude would be required for other boiler sizes and emission limits. Of course, if detailed design and cost calculations were performed so as to reduce the uncertainty of the cost comparisons, clear economic choices between the three technology options could be made on a case-by-case basis.

6.4 Coal Price Sensitivity

Since fuel changes represent a significant portion of the total annual costs for each of the SO_2 control alternatives examined, it is useful to quantify the impact of coal price changes on model boiler total annual costs. The algorithm format of the total annual cost estimation procedure allows ready derivation of formulas for coal price sensitivity. These formulas are presented in Table 6.5-1 for the model boilers examined in this study.

Annual costs for a 150 million Btu/hr boiler operated to meet a 1.2 lb $SO_2/10^6$ Btu emission limit are used to illustrate the coal price sensitivity of the various SO_2 technology alternatives. Using the formulas from the table, one can show that a $1.00/million Btu coal price increase translates to an annual cost increase of $795,000 for an FBC boiler and $788,000 for a spreader stoker boiler equipped with LEA NO_x control. The latter cost increase applies equally to both the compliance coal and the FGD control alternatives. For a pulverized coal boiler equipped with LEA/SCA NO_x

TABLE 6.5-1. COAL PRICE SENSITIVITY OF TOTAL ANNUAL COSTS FOR MODEL BOILERS

For an FBC boiler:

$$\Delta TAC = 8833 \times CF \times Q \times \Delta FC$$

For a spreader stoker boiler (with LEA NO_x control):

$$\Delta TAC = CF \times Q \times \Delta FC\ [8833 - 5.5 \times 10^{-4} \times FFAC \times (UNCEA - CTREA)]$$

For a pulverized coal boiler (with LEA/SCA NO_x control):

$$\Delta TAC = 8855 \times CF \times Q \times \Delta FC$$

Where,

TAC = Total annual costs, \$.
CF = Capacity factor, expressed as a decimal.
Q = Boiler heat input capacity, 10^6 Btu/hr.
FC = Fuel cost, \$/10^6 Btu.
FFAC = F factor, Dry SCF/10^6 Btu heat input (9820 for coal).
UNCEA = Uncontrolled excess air, percent.
CTREA = Controlled excess air, percent.

control, the annual cost change due to a \$1.00/million Btu coal price change is \$797,000. Again, this increase applies equally to both the compliance coal and FGD alternatives. The nearness of the total annual cost changes indicates that the coal price sensitivities of the three SO_2 control alternatives are equivalent for practical purposes.

6.5 CONCLUSIONS

The overall conclusion that can be drawn from the cost data and analysis of this section is that annual cost differences among FBC technology, conventional boiler/FGD systems, and compliance coal combustion are expected to be small (± 15 percent or less) over the range of SO_2 emission limitations and boiler sizes examined. Absolute economic competitiveness among these alternatives will be determined by site-specific parameters. In addition, FBC cost data show that Ca/S ratios have only a minor effect on system capital and operating costs; significant reductions in the required Ca/S ratio for a given level of SO_2 removal (which is an objective of research at the Tennessee Valley Authority pilot plant and elsewhere) will not noticeably alter the economic competitiveness of FBC technology for industrial applications.

Given the small cost differences among the studied SO_2 control alternatives in the current context, and the lack of expectations for dramatic changes in the near future, it is unlikely that economics alone will be the deciding factor when a choice is made among options by an industrial plant owner. Rather, less tangible factors such as requirements for fuel flexibility and preference for risk are likely to play more important roles in the decision process.

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APPENDIX A

FBC COST ALGORITHM DEVELOPMENT

In 1979, the FBC-ITAR cost estimates were translated into cost algorithms by Acurex Corporation.[1] The Acurex algorithms are generally faithful to the ITAR design basis and costs. Exceptions were noted on review, however, and were corrected as summarized below:

- The Acurex expressions for turnkey costs for limestone and spent solids storage and handling seriously underestimated the ITAR costs. These expressions were revised to duplicate the original estimation procedures outlined by GCA in the ITAR;

- The term for supervisory labor had been left out of the expression for plant overhead costs; this oversight was corrected.

- A correlation had been developed for flue gas flow rate as a function boiler size but data for air flow rates to the boiler had been used instead of flue gas rates. A new expression for flue gas flow was derived from the flue gas rate-versus-boiler capacity data in Table C-5 of the ITAR;

In addition, a number of algorithm modifications were made to make the final expressions consistent with existing algorithms for conventional boilers and air pollution control devices and/or more flexible for use in this study. These modifications included:

- Added provisions for estimating costs for a 400 million Btu/hr boiler. The largest boiler which had been costed in the ITAR was a 200 million Btu/hr unit. A recent study by Combustion Engineering, Inc. indicates that 250 million Btu/hr is the maximum capacity for shop-assembled, rail-shippable FBC boilers.[2] However, the ITAR costs were based on a 30 million Btu/hr fully shop fabricated unit; a 75 million Btu/hr unit that was field erected from shop fabricated modules; and fully field erected 150 and 200 million Btu/hr units. Since the ITAR cost basis did not extend to a 400 million Btu/hr unit, two 200 million Btu/hr FBC boilers were specified for the 400 million Btu/hr case. This unit has a single train of limestone and spent solids storage and handling equipment, however. Appropriate factors were applied to capital cost estimates as recommended by PEDCo for dual unit boilers;[3]

- Eliminated Acurex equations which predicted Ca/S ratio as a function of SO_2 removal efficiency. In this report, the Ca/S ratios used in cost calculations are those projected by the Westinghouse model as summarized in Table 5.1-2 (or extrapolated via power curve). To provide greater flexibility, Ca/S ratios are now specified as input data by the user;

- Added an expression to calculate uncontrolled particulate matter from the FBC unit. The FBC boiler design includes a primary cyclone for solids recycle. To maintain consistency with the ITAR, the flow of PM from the cyclone was set equal to 10 percent of the non-combustible solids flow (i.e., coal ash, unreacted limestone, calcined limestone, and sulfated limestone) into the boiler. This ratio was selected in the ITAR because it was consistent with the experimentally documented range of particulate matter loadings at the primary cyclone exit. Based on ITAR mass flow rates, the solids recycle rate varies from 0.2 to 0.4. The algorithm expression incorporates this range of recycle rates;

- Revised the expression for working capital to be consistent with algorithms for other technologies (see Appendix D);

- Adjusted the costs for performance tests from $12,000 in the Acurex algorithms to 1 percent of boiler total direct cost; this specification is consistent with other algorithms (see Appendix D);

- Added a labor factor to these same equations to account for reduced labor requirements at reduced capacity to maintain consistency with other algorithms (see Appendix D);

- Added provisions to revise capital and annual costs to a different time basis using capital equipment cost indices and specific unit costs;

The resulting cost algorithm for industrial atmospheric FBC technology is listed in Table A-1. A description of the terms used in the algorithm and their corresponding units are contained in Table A-2.

TABLE A-1. COST EQUATIONS FOR COAL-FIRED FLUIDIZED BED COMBUSTION (FBC) BOILERS[a]

Routine Code:

Capital Costs:

$TK = TKB + TKLS + TKSW$

$TKB = 1.596 * TDB \qquad Q \leq 58.6\ MW$

$ = 1.484 * TDB \qquad Q > 58.6\ MW$

where

$TDB = (814{,}200 + 362{,}000\ (Q - 8.8)^{0.7})\left(1.23 - \frac{8.21\ H}{10^6}\right) \text{for } Q \geq 58.6\ MW$

$TDB = 1.748\ (814{,}200 + 361{,}000\ (Q/2 - 8.8)^{0.7})\left(1.23 - \frac{8.21H}{10^6}\right) Q > 73.2\ MW$

$TKLS = 2.317\ (CL * VCL + 4.4 * LSFR)$

$CL = 0.2409 * LSFR$

$VCL = 349.3 - 0.244\ CL \qquad CL \leq 283$

$VCL = 383 \qquad CL > 283$

$LSFR = (Q/H)\ (1.25 \times 10^5)\ (S)\ (FCS)$

$TKSW = 2.422 * CW * VCW$

$CW = 0.2139 * SWFR$

$VCW = 396.8 - 0.3278\ CW \qquad CW \leq 283$

$VCW = 431 \qquad CW > 283$

$SWFR = 0.9\ (0.624 * LSFR + CFR\left(\frac{(S)(EFFSO_2(2.5)}{10{,}000} + \frac{A}{100}\right)\Big)$

$CFR = 3.6 \times 10^6\ (Q/H)$

$TD = TDB + \frac{TKLS}{1.56} + \frac{TKSW}{1.56}$

TABLE A-1. COST EQUATIONS FOR COAL-FIRED FLUIDIZED BED COMBUSTION (FBC) BOILERS[a] (Continued)

$\mathrm{IND} = 0.33\ \mathrm{TDB} + \dfrac{0.3(\mathrm{TKLS} + \mathrm{TKSW})}{1.56}$	$Q \geq 58.6$
$\mathrm{IND} = 0.237\ \mathrm{TDB} + \dfrac{0.3\ (\mathrm{TKLS} + \mathrm{TKSW})}{1.56}$	$Q \geq 58.6$
Annual Costs	
$\mathrm{DL} = \mathrm{LF} * 123{,}000\ \mathrm{Exp}\ (0.02 * Q)\ (\mathrm{DLR}/12.02)$	$Q \leq 58.6$
$\mathrm{DL} = \mathrm{LF} * 397{,}100\ (\mathrm{DLR}/12.02)$	$Q > 58.6$
$\mathrm{SPRV} = \mathrm{LF} * 62{,}520 * (\mathrm{SLR}/15.63)$	$Q < 15$
$= \mathrm{LF} * 125{,}040 * (\mathrm{SLR}/15.63)$	$Q > 15$
$\mathrm{MANT} = 58{,}500 * \mathrm{LF} * (\mathrm{AMLR}/14.63)$	$Q \leq 15$
$= 117{,}000 * \mathrm{LF} * (\mathrm{AMLR}/14.63)$	$15 < Q \leq 50$
$= 176{,}000 * \mathrm{LF} * (\mathrm{AMLR}/14.63)$	$15 < Q$
$\mathrm{SP} = 157{,}000\ \mathrm{EXP}\ (2.52 \times 10^{-7}\ (\mathrm{TDB}) - 3.8 \times 10^{15}\ (H))$	
$\mathrm{ELEC} = 8{,}760\ (\mathrm{CF})\ (\mathrm{ELECR})\ (19.82\ Q - 1.78)$	
$\mathrm{WT} = 8{,}760\ (\mathrm{CF})\ (\mathrm{WTRR})\ (2.06)\ (Q)$	
$\mathrm{FUEL} = 8{,}760\ (\mathrm{CF})\ (\mathrm{FC})\ (Q)\ (3{,}600)$	
$\mathrm{LMS} = 8{,}760\ (\mathrm{CF})\ (\mathrm{LSFR}) * (\mathrm{ALS})$	
$\mathrm{SW} = 8{,}760\ (\mathrm{CF})\ (\mathrm{SWDR})\ (0.9)\left(\dfrac{0.624\ \mathrm{LMS}}{\mathrm{LSFR}} + \dfrac{\mathrm{FUEL}}{\mathrm{FC}}\left(\dfrac{(2.5\ (\mathrm{EFFSO_2})\ (S)}{10{,}000} + \dfrac{A}{100}\right)\right)$	

A conservative estimate of FCS is:

$$\mathrm{FCS} = 7.605 \times 10^{-5}\ \mathrm{EFFSO_2}\ 2.431$$

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[a] FBC algorithm uses metric units as shown in Table A-2.

TABLE A-2. NOMENCLATURE FOR FBC ALGORITHM

Term	Description
A	Ash content (wt. percent)
ALS	Limestone Rate ($/hr)
AMLR	Maintenance Labor Rate ($/man-hr)
CF	Capacity Factor (unit less)
CFR	Coal Feed Rate (kg/hr)
CL	Limestone Storage Capacity (m^3)
CW	Solid Waste Storage Capacity (m^3)
DLR	Direct Labor Rate ($/man-hr)
$EFFSO_2$	SO_2 Removal Efficiency (percent)
ELECR	Electricity Rate ($/kw-hr)
FC	Fuel Cost ($\$/10^6$ Btu)
FCS	Calcium to Sulfur Ratio (unit less)
FUEL	Annual Fuel Cost ($/year)
H	Heating Value (Btu/lb)
LF	Labor Factor (unit less)
LMS	Annual Limestone Cost ($/year)
LSFR	Limestone Feed Rate
Q	Heat Input (10^6 Btu/hr)
S	Sulfur Content (wt. percent)
SLR	Supervision Labor Rate ($/man-hr)
SWDR	Solid Waste Rate ($/kg)
SWFR	Solid Waste Feed Rate (kg/hr)
TDB	Total Direct Boiler Cost ($)
TKB	Boiler Turnkey Cost ($)
TKLS	Limestone Turnkey Cost ($)
TKSW	Solid Waste Turnkey Cost ($)
VCL	Limestone Storage Cost ($\$/m^3$)
VCW	Solid Waste Storage Cost ($\$/m^3$)

APPENDIX A REFERENCES

1. Gardner, R., R. Chang, and L. Broz. (Acurex Corporation.) Cost, Energy and Environmental Algorithms for NO_x, SO_2 and PM controls for Industrial Boilers. Final Report. (Prepared for U. S. Environmental Protection Agency.) Cincinnati, Ohio. EPA Contract No. 68-03-2567. December 1979. p. 20-52.

2. Myrick, D. T. (Combustion Engineering, Inc.) DOE Cost Comparison Study: Industrial Fluidized Bed Combustion Vs. Conventional Coal Technology. (Prepared for U. S. Department of Energy.) FE-2473-T7. January 1980.

3. Devitt, T., P. Spaite, and L. Gibbs. (PEDCo Environmental) Population and Characteristics of Industrial/Commercial Boilers in the U. S. (Prepared for U. S. Environmental Protection Agency.) Research Triangle Park, N. C. EPA-600/7-79-78a. Cincinnati, Ohio. August 1979. 462 p.

APPENDIX B

SUMMARY OF CAPITAL AND OPERATING COSTS FOR MODEL BOILERS

Model boiler costs for the three SO_2 control limits examined in this study are summarized in this appendix. Costs are segregated by boiler, NO_x control, SO_2 control, and PM control equipment and normalized on the basis of boiler heat input capacity.

TABLE B-1. CAPITAL COSTS OF MODEL BOILERS FOR SO_2 STANDARD = 1.7 LB/10^6 BTU
(JANUARY 1983, DOLLARS)

Control Alternative	Model Boiler	Capital Costs ($1000) Boiler	NO_x Control[d]	SO_2 Control[e]	PM[f] Control	Total	Normalized Total $\left(\frac{\$1000}{10^6\ \text{Btu/hr}}\right)$
1A	50-FBC, Type H[g], 75, 2.75, FF	5,273	-	-	477	5,750	115
1B	50-FBC, Type H, 75, 1.85, FF	5,194	-	-	477	5,671	113
2	50-FGD, Type H, 75, LEA[b]	3,716	19	1,368	-	5,103	102
3	50-CC, Type D, FF, LEA[c]	3,515	19	-	594	4,128	83
1A	100-FBC, Type H, 75, 2.75, FF	9,823	-	-	922	10,745	107
1B	100-FBC, Type H, 75, 1.85, FF	9,596	-	-	921	10,518	105
2	100-FGD, Type H, 75 LEA	7,924	24	1,994	-	9,942	99
3	100-CC, Type D, 55, LEA	7,760	24	-	1,090	8,874	89
1A	150-FBC, Type H, 75, 2.75, FF	13,656	-	-	1,273	14,929	100
1B	150-FBC, Type H, 75, 1.85, FF	13,345	-	-	1,272	14,616	98
2	150-FGD, Type H, 75 LEA	11,110	29	2,498	-	13,637	91
3	150-CC, Type D, FF, LEA	10,883	29	-	1,489	12,401	83

[a] Boiler size-technology, coal type, SO_2 removal (percent), Ca/S ratio, PM control device.

[b] Boiler size-technology, coal type, SO_2 removal (percent), NO_x control technique.

[c] Boiler size-technology, coal type, PM control device, NO_x control technique.

[d] NO_x control intrinsic to FBC boiler.

[e] SO_2 control intrinsic to FBC boiler.

[f] PM control intrinsic to lime spray drying FGD system.

[g] All coal types are bituminous coals except where noted.

TABLE B-2. CAPITAL COSTS OF MODEL BOILERS FOR SO_2 CONTROL - 1.2 LB/10^6 BTU
(JANUARY 1983, DOLLARS)

Control Alternative	Model Boiler	Capital Costs ($1000) Boiler	NO_x Control[d]	SO_2 Control[e]	PM[f] Control	Total	Normalized Total $\left(\frac{\$1000}{10^6\ \text{Btu/hr}}\right)$
1A	50-FBC, Type H, 80, 3.2, FF	5,313	-	-	477	5,790	116
1B	50-FBC, Type H, 80, 2.2, FF	5,227	-	-	477	5,704	114
1C	50-FBC, Type F, 65, 1.95, FF	5,123	-	-	476	5,599	112
1D	50-FBC, Type F, 65, 1.25, FF	5,087	-	-	476	5,564	111
2A	50-FGD, Type H, 80, LEA[b]	3,716	19	1,400	-	5,135	103
2B	50-FGD, Type F, 65, LEA	3,786	19	1,167	-	4,969	99
3A	50-CC, Type B, FF, LEA[c]	4,831	19	-	623	5,473	109
3B	50-CC, Type B, FF, LEA[h]	3,814	19	-	594	4,427	89
1A	100-FBC, Type H, 80, 3.2, FF	10,054	-	-	922	10,976	110
1B	100-FBC, Type H, 80, 2.2, FF	9,651	-	-	922	10,573	106
1C	100-FBC, Type F, 65, 1.95 FF	9,479	-	-	920	10,400	104
1D	100-FBC, Type F, 65, 1.25, FF	9,414	-	-	920	10,334	103
2A	100-FGD, Type H, 80, LEA	7,924	24	2,041	-	9,989	100
2B	100-FGD, Type F, 65, LEA	7,991	24	1,713	-	9,728	97
3A	100-CC, Type B, FF, LEA	8,737	24	-	1,134	9,895	99
3B	100-CC, Type B, FF, LEA[h]	8,006	24	-	1,090	9,120	91
1A	150-FBC, Type H, 80, 3.2, FF	13,819	-	-	1,273	15,092	101
5B	150-FBC, Type H, 80, 2.2, FF	13,473	-	-	1,272	14,745	98
1C	150-FBC, Type F, 65, 1.95, FF	12,925	-	-	1,270	14,195	95
1D	150-FBC, Type F, 65, 1.25, FF	12,836	-	-	1,270	14,106	94
2A	150-FGD, Type H, 80, LEA	11,110	30	2,559	-	13,699	91

TABLE B-2. (CONTINUED) CAPITAL COSTS OF MODEL BOILERS FOR SO_2 CONTROL = 1.2 LB/10^6 BTU
(JANUARY 1983, DOLLARS)

		Capital Costs ($1000)					
Control Alternative	Model Boiler	Boiler	NO_x Control[d]	SO_2 Control[e]	PM[f] Control	Total	Normalized Total $\left(\frac{\$1000}{10^6 \text{ Btu/hr}}\right)$
2B	150-FGD, Type F, 65, LEA	11,206	29	2,144	-	13,379	89
3A	150-CC, Type B, FF, LEA	12,253	29	1,546	-	13,828	92
3B	150-CC, Type B, FF, LEA[h]	11,228	29	-	-	12,746	85
1A	250-FBC, Type H, 80, 3.2, FF	20,373	-	-	1,870	22,243	89
1B	250-FBC, Type H, 80, 2.2, FF	19,797	-	-	1,869	21,666	87
1C	250-FBC, Type F, 65, 1.95, FF	19,002	-	-	1,865	20,867	83
1D	250-FBC, Type F, 65, 1.25, FF	18,837	-	-	1,865	20,702	83
2A	250-FGD, Type H, 80, SCA	19,101	89	3,676	-	22,766	91
2B	250-FGD, Type F, 65, SCA	19,218	89	2,975	-	22,282	89
3A	250-CC, Type B, FF, SCA	19,979	89	2,201	-	22,269	89
3B	250-CC, Type B, FF, SCA[h]	18,905	89	2,118	-	21,113	84
1A	400-FBC, Type H, 80, 3.2, FF	29,024	-	-	2,655	31,679	79
1B	400-FBC, Type H, 80, 2.2, FF	28,102	-	-	2,652	30,754	77
1C	400-FBC, Type F, 65, 1.95, FF	26,957	-	-	2,647	29,604	74
1D	400-FBC, Type F, 65, 1.25, FF	26,534	-	-	2,646	29,179	73
2A	400-FGD, Type H, 80, SCA	26,341	127	4,856	-	31,324	78
2B	400-FGD, Type F, 65, SCA	26,502	127	4,056	-	30,685	77
3A	400-CC, Type B, FF, SCA	27,403	127	-	3,129	30,659	77
3B	400-CC, Type B, FF, SCA[h]	26,144	127	-	3,010	29,281	73

TABLE B-3. CAPITAL COSTS OF MODEL BOILERS FOR SO_2 STANDARD = 0.8 LB/10^6 BTU
(JANUARY 1983, DOLLARS)

		Capital Costs ($1000)					
Control Alternative	Model Boiler	Boiler	NO_x Control[d]	SO_2 Control[e]	PM[f] Control	Total	Normalized Total $\left(\frac{\$1000}{10^6\ Btu/hr}\right)$
1A	50-FBC, Type H, 90, 4.3, FF	5,404	-	-	478	5,881	118
1B	50-FBC, Type H, 90, 2.8, FF	5,283	-	-	477	5,760	115
1C	50-FBC, Type F, 80, 3.2, FF	5,187	-	-	476	5,664	113
1D	50-FBC, Type F, 80, 2.2, FF	5,138	-	-	476	5,615	112
2A	50-FGD, Type H, 90, LEA[b]	3,716	19	1,480	-	5,215	104
2B	50-FGD, Type F, 80, LEA	3,786	19	1,226	-	5,028	101
3A	50-CC, Type A, FF, LEA[c]	4,831	19	-	623	5,473	109
3B	50-CC, Type B, FF, LEA[h]	3,545	19	-	594	4,157	83
1A	100-FBC, Type H, 90, 4.3, FF	10,317	-	-	923	11,240	112
1B	100-FBC, Type H, 90, 2.8, FF	9,971	-	-	922	10,893	109
1C	100-FBC, Type F, 80, 3.2, FF	9,593	-	-	921	10,514	105
1D	100-FBC, Type F, 80, 2.2, FF	9,507	-	-	921	10,428	104
2A	100-FGD, Type H, 90, LEA	7,924	24	2,163	-	10,111	101
2B	100-FGD, Type F, 80, LEA	7,991	24	1,797	-	9,812	98
3A	100-CC, Type A, FF, LEA	8,737	24	-	1,134	9,895	99
3B	100-CC, Type A, FF, LEA[h]	8,013	24	-	1,090	9,127	91
1A	150-FBC, Type H, 90, 4.3, FF	14,214	-	-	1,275	15,488	103
1B	150-FBC, Type H, 90, 2.8, FF	13,695	-	-	1,273	14,968	100
1C	150-FBC, Type F, 80, 3.2, FF	13,212	-	-	1,271	14,483	96
1D	150-FBC, Type F, 80, 2.2, FF	12,962	-	-	1,271	14,232	95
2A	150-FGD, Type H, 90, LEA	11,110	30	2,717	-	13,857	92

TABLE B-3. (CONTINUED) CAPITAL COSTS OF MODEL BOILERS FOR SO_2 STANDARD - 0.8 LB/10^6 BTU
(JANUARY 1983, DOLLARS)

Control Alternative	Model Boiler	Capital Costs ($1000)					
		Boiler[d]	NO_x Control[e]	SO_2 Control	PM[f] Control	Total	Normalized Total $\left(\frac{\$1000}{10^6\ \text{Btu/hr}}\right)$
2B	150-FGD, Type F, 80, LEA	11,206	29	2,251	-	13,486	90
3A	150-CC, Type A, FF, LEA	12,253	29	-	1,546	13,828	92
3B	150-CC, Type A, FF, LEA[h]	11,239	29	-	1,409	12,757	85
1A	250-FBC, Type H, 90, 4.3, FF	21,031	-	-	1,873	22,904	92
1B	250-FBC, Type H, 90, 2.8, FF	20,167	-	-	1,870	22,037	88
1C	250-FBC, Type F, 80, 3.2, FF	19,479	-	-	1,867	21,346	85
1D	250-FBC, Type F, 80, 2.2, FF	19,181	-	-	1,866	21,047	84
2A	250-FGD, Type H, 90, SCA	19,101	89	3,807	-	22,997	92
2B	250-FGD, Type F, 80, SCA	19,218	89	3,125	-	22,432	90
3A	250-CC, Type A, FF, SCA	19,979	89	-	2,201	22,269	89
3B	250-CC, Type B, FF, SCA[h]	18,923	89	-	2,118	21,130	85
1A	400-FBC, Type H, 90, 4.3, FF	30,077	-	-	2,658	32,735	82
1B	400-FBC, Type H, 90, 2.8, FF	28,694	-	-	2,654	31,348	78
1C	400-FBC, Type F, 80, 3.2, FF	27,582	-	-	2,649	30,231	76
1D	400-FBC, Type F, 80, 2.2, FF	27,106	-	-	2,647	29,754	74
2A	400-FGD, Type H, 90, SCA	26,341	127	5,175	-	31,643	79
2B	400-FGD, Type F, 80, SCA	26,502	127	4,258	-	30,887	77
3A	400-CC, Type A, FF, SCA	27,403	127	-	3,129	30,659	77
3B	400-CC, Type A, FF, SCA[h]	26,172	127	-	3,010	29,309	73

TABLE B-4. ANNUAL COSTS OF MODEL BOILERS FOR SO_2 STANDARD = 1.7 LB/10^6 BTU
(JANUARY 1983, DOLLARS)

Control Alternative	Model Boiler	Annual Costs ($1000)					Normalized Total $/$10^6$ Btu
		Boiler	NO_x Control[d]	SO_2 Control[e]	PM[f] Control	Total	
1A	50-FBC, Type H, 75, 2.75, FF	2,139	-	-	139	2,278	8.4
1B	50-FBC, Type H, 75, 1.85, FF	2,105	-	-	137	2,243	8.3
2	50-FGD, Type H, 75, LEA[b]	1,778	-2	506	-	2,282	8.7
3	50-CC, Type D, FF, LEA[c]	1,923	-4	-	157	2,076	7.6
1A	100-FBC, Type H, 75, 2.75, FF	3,981	-	-	247	4,228	8.1
1B	100-FBC, Type H, 75, 1.85, FF	3,901	-	-	244	4,145	7.9
2	100-FGD, Type H, 75 LEA	3,299	-6	726	-	4,019	7.6
3	100-CC, Type D, FF, LEA	3,661	-10	-	280	3,931	7.1
1A	150-FBC, Type H, 75, 2.75, FF	5,622	-	-	339	5,961	7.6
1B	150-FBC, Type H, 75, 1.85, FF	5,507	-	-	335	5,841	7.4
2	150-FGD, Type H, 75 LEA	4,649	-11	916	0	5,554	7.0
3	150-CC, Type D, FF, LEA	5,196	-16	-	382	5,562	7.1

TABLE B-5. ANNUAL COSTS OF MODEL BOILERS FOR SO_2 STANDARD = 1.2 lb/10^6 Btu (JANUARY 1983, DOLLARS)

Control Alternative	Model Boiler	Annual Costs ($1000) Boiler	NO_x Control[d]	SO_2 Control[e]	PM[f] Control	Total	Normalized Total $/$10^6$ Btu
1A	50-FBC, Type H, 80, 3.2, FF	2,157	-	-	140	2,297	8.7
1B	50-FBC, Type H, 80, 2.2, FF	2,120	-	-	138	2,258	8.6
1C	50-FBC, Type F, 65, 1.95, FF	2,191	-	-	135	2,326	8.9
1D	50-FBC, Type F, 65, 1.25, FF	2,177	-	-	135	2,312	8.8
2A	50-FGD, Type H, 80, LEA[b]	1,778	-2	525	-	2,301	8.8
2B	50-FGD, Type F, 65, LEA	1,912	-3	421	-	2,330	8.9
3A	50-CC, Type B, FF, LEA[c]	2,105	-3	-	164	2,266	8.6
3B	50-CC, Type B, FF, LEA[b]	2,007	-4	-	157	2,160	8.2
1A	100-FBC, Type H, 80, 3.2, FF	4,043	-	-	248	4,291	8.2
1B	100-FBC, Type H, 80, 2.2, FF	3,928	-	-	245	4,173	7.9
1C	100-FBC, Type F, 65, 1.95, FF	4,076	-	-	240	4,316	8.2
1D	100-FBC, Type F, 65, 1.25, FF	4,049	-	-	239	4,287	8.2
2A	100-FGD, Type H, 80, LEA	3,299	-6	760	-	4,053	7.7
2B	100-FGD, Type F, 65, LEA	3,555	-9	578	-	4,124	7.8
3A	100-CC, Type B, FF, LEA	3,629	-8	-	294	3,915	7.4
3B	100-CC, Type B, FF, LEA[b]	3,735	-11	-	280	4,004	7.6
1A	150-FBC, Type H, 80, 3.2, FF	5,683	-	-	341	6,024	7.6
1B	150-FBC, Type H, 80, 2.2, FF	5,555	-	-	336	5,891	7.5
1C	150-FBC, Type F, 65, 1.95, FF	5,728	-	-	328	6,056	7.7
1D	150-FBC, Type F, 65, 1.25, FF	5,688	-	-	327	6,014	7.6
2A	150-FGD, Type H, 80, LEA	4,649	-11	966	-	5,604	7.1

TABLE B-5. (CONTINUED) ANNUAL COSTS OF MODEL BOILERS FOR SO_2 STANDARD = 1.2 LB/10^6 BTU
(JANUARY 1983, DOLLARS)

Control Alternative	Model Boiler	Annual Costs ($1000)					Normalized Total
		Boiler	NO_x Control[d]	SO_2 Control[e]	PM[f] Control	Total	$/$10^6$ Btu
2B	150-FGD, Type F, 65, LEA	5,033	-14	708	-	5,727	7.3
3A	150-CC, Type B, FF, LEA	5,131	-14	-	402	5,319	7.0
3B	150-CC, Type B, FF, LEA[h]	5,302	-17	-	382	5,667	7.2
1A	250-FBC, Type H, 80, 3.2, FF	9,001	-	-	509	9,510	7.2
1B	250-FBC, Type H, 80, 2.2, FF	8,787	-	-	502	9,289	7.1
1C	250-FBC, Type F, 65, 1.95, FF	9,097	-	-	488	9,586	7.3
1D	250-FBC, Type F, 65, 1.25, FF	9,028	-	-	486	9,513	7.2
2A	250-FGD, Type H, 80, SCA	8,098	58	1,388	-	9,504	7.2
2B	250-FGD, Type F, 65, SCA	8,691	60	972	-	9,723	7.4
3A	250-CC, Type B, FF, SCA	8,697	60	-	585	9,332	7.1
3B	250-CC, Type B, FF, SCA[h]	9,080	-	-	570	9,709	7.4
1A	400-FBC, Type H, 80, 3.2, FF	14,548	-	-	745	15,293	7.3
1B	400-FBC, Type H, 80, 2.2, FF	14,206	-	-	733	14,939	7.1
1C	400-FBC, Type F, 65, 1.95, FF	14,740	-	-	712	15,451	7.3
1D	400-FBC, Type F, 65, 1.25, FF	14,600	-	-	708	15,308	7.3
2A	400-FGD, Type H, 80, SCA	11,761	90	1,959	-	13,810	6.6
2B	400-FGD, Type F, 65, SCA	12,768	92	1,323	-	14,183	6.7
3A	400-CC, Type B, FF, SCA	12,708	92	-	856	13,656	6.5
3B	400-CC, Type B, FF, SCA[h]	13,415	-	-	835	14,342	6.8

TABLE B-6. ANNUAL COSTS OF MODEL BOILERS FOR SO_2 STANDARD = 0.8 LB/10^6 BTU
(JANUARY 1983, DOLLARS)

Control Alternative	Model Boiler	Annual Costs ($1000) Boiler	NO_x Control[d]	SO_2 Control[e]	PM[f] Control	Total	Normalized Total $/$10^6$ Btu
1A	50-FBC, Type H, 90, 4.3, FF	2,200	-	-	141	2,341	8.9
1B	50-FBC, Type H, 90, 2.8, FF	2,144	-	-	139	2,284	8.7
1C	50-FBC, Type F, 80, 3.2, FF	2,218	-	-	136	2,355	9.0
1D	50-FBC, Type F, 80, 2.2, FF	2,198	-	-	136	2,334	8.9
2A	50-FGD, Type H, 90, LEA[b]	1,778	-2	579	-	2,355	9.0
2B	50-FGD, Type F, 80, LEA	1,912	-3	449	-	2,358	0.9
3A	50-CC, Type A, FF, LEA[c]	2,105	-3	-	164	2,266	8.3
3B	50-CC, Typu A, FF, LEA[h]	1,987	-4	-	157	2,140	8.1
1A	100-FBC, Type H, 90, 4.3, FF	4,142	-	-	252	4,393	8.4
1B	100-FBC, Type H, 90, 2.8, FF	4,013	-	-	247	4,261	8.1
1C	100-FBC, Type F, 80, 3.2, FF	4,128	-	-	242	4,370	8.3
1D	100-FBC, Type F, 80, 2.2, FF	4,089	-	-	240	4,330	8.3
2A	100-FGD, Type H, 90, LEA	3,299	-6	861	-	4,154	7.9
2B	100-FGD, Type F, 80, LEA	3,555	-9	627	-	4,173	7.9
3A	100-CC, Type A, FF, LEA	3,629	-8	-	294	3,915	7.4
3B	100-CC, Type A, FF, LEA[h]	3,820	-11	-	280	4,088	7.8
1A	150-FBC, Type H, 90, 4.3, FF	5,831	-	-	346	6,177	7.8
1B	150-FBC, Type H, 90, 2.8, FF	5,639	-	-	340	5,978	7.6
1C	150-FBC, Type F, 80, 3.2, FF	5,827	-	-	332	6,159	7.8
1D	150-FBC, Type F, 80, 2.2, FF	5,746	-	-	329	6,076	7.7
2A	150-FGD, Type H, 90, LEA	4,649	-11	1,113	-	5,751	7.3

TABLE B-6. (CONTINUED) ANNUAL COSTS OF MODEL BOILERS FOR SO_2 STANDARD = 0.8 LB/10^6 BTU
(JANUARY 1983, DOLLARS)

Control Alternative	Model Boiler	Annual Costs ($1000)					Normalized Total $/$10^6$ Btu
		Boiler	NO_x Control^d	SO_2 Control^e	PM^f Control	Total	
2B	150-FGD, Type F, 80, LEA	5,033	-14	778	-	5,797	7.4
3A	150-CC, Type A, FF, LEA	5,131	-14	-	402	5,619	7.0
3B	150-CC, Type A, FF, LEA^h	5,429	-19	-	382	5,793	7.3
1A	250-FBC, Type H, 90, 4.3, FF	9,248	-	-	518	9,765	7.4
1B	250-FBC, Type H, 90, 2.8, FF	8,927	-	-	507	9,434	7.2
1C	250-FBC, Type F, 80, 3.2, FF	9,259	-	-	494	9,753	7.4
1D	250-FBC, Type F, 80, 2.2, FF	9,149	-	-	490	9,638	7.3
2A	250-FGD, Type H, 90, SCA	8,058	58	1,627	-	9,743	7.4
2B	250-FGD, Type F, 80, SCA	8,691	60	1,083	-	9,834	7.5
3A	250-CC, Type A, FF, SCA	8,687	60	-	585	9,332	7.1
3B	250-CC, Type A, FF, SCA^h	9,292	60	-	570	9,922	7.6
1A	400-FBC, Type H, 90, 4.3, FF	14,943	-	-	759	15,702	7.5
1B	400-FBC, Type H, 90, 2.8, FF	14,429	-	-	742	15,171	7.2
1C	400-FBC, Type F, 80, 3.2, FF	14,975	-	-	720	15,695	7.5
1D	400-FBC, Type F, 80, 2.2, FF	14,798	-	-	714	15,512	7.4
2A	400-FGD, Type H, 90, SCA	11,761	90	2,332	-	14,183	6.7
2B	400-FGD, Type F, 80, SCA	12,768	92	1,494	-	14,354	6.8
3A	400-CC, Type A, FF, SCA	12,708	92	-	3,119	15,929	6.5
3B	400-CC, Type A, FF, SCA^h	13,754	93	-	835	14,682	7.0

APPENDIX C
ADJUSTMENTS TO INDEPENDENT COST ESIMTATES

This appendix summarizes details of the adjustments that have been made to FBC cost estimates developed by independent workers. The purpose of the adjustments was to place all estimates on a common design and scope basis so that fair comparisons can be made among them.

C.1 COMBUSTION ENGINEERING, INC. ESTIMATE

This estimate is derived from a report which projects costs for a new FBC boiler located in Ft. Wayne, Indiana producing 250,000 lb/hr steam at 900 psig and 750°F.[1] Two FBC designs are considered in this study: (1) Two shop assembled, rail-shippable units rated at 125,000 lb/hr, and (2) a single field assembled unit producing 250,000 lb/hr steam. Since the FBC algorithm specifies dual boilers for this size (352 million Btu/hr input), the first case was selected for comparison. The CE estimate is based on detailed equipment designs and layout and internal cost files. Other important factors in the CE system design include:

- Air emission standards:
    - 1.2 lb $SO_2/10^6$ Btu plus 85 percent reduction
    - 0.5 lb $NO_x/10^6$ Btu
    - 0.03 lb $PM/10^6$ Btu plus 99 percent reduction
- Coal: Midwest bituminous, 10,430 Btu/lb, 3.5 percent sulfur, 9.2 percent ash
- Coal and limestone handling:
    - Coal - crushing, drying, and 2 days prepared coal storage

- Limestone: 4 days storage of crushed and sized limestone, 1/8 inch particle size
- Solid waste disposal: landfilled at a site adjacent to the plant, 6 days on-site storage
- Ca/S Ratio: 3.0
- Mid-1979 cost basis
- Load factor of 0.68
- Boiler efficiency of 84 percent

Table C-1 shows the major adjustments made to the CE estimates to achieve compatibility with FBC algorithm projections. These adjustments resulted in a total capital cost of $37,473,000 and a total annual cost of $13,268,000/year.

C-2 FOSTER WHEELER ESTIMATE

The FW estimate corresponds to new industrial FBC boiler generating 212,000 lb/hr of steam at 650 psig and 750°F.[2] Costs are estimated for both Western and Eastern coal operation; only the Eastern coal costs are presented here. The specified coal feed rate and heat content correspond to 291 million Btu/hr heat input. The FW estimate was developed from detailed equipment designs and internal cost files. Other particulars of the FW estimate include:

- Air emission standards:
  - 1.2 lb $SO_2/10^6$ Btu
  - 0.5 lb $NO_x/10^6$ Btu
  - 0.03 lb $PM/10^6$ Btu

TABLE C-1. MAJOR ADJUSTMENTS TO THE COMBUSTION ENGINEERING COST BASIS

1. Contingencies on new product design were subtracted from total delivered capital costs; re-estimated at 20 p9ercent of direct plus indirect costs.
2. Land costs (for landfill adjacent to boiler site) were subtracted except for $6000.
3. A load factor of 0.6 (as opposed to 0.68) was used to determine annual costs; a labor factor of 0.75 was applied.
4. Capital costs were updated from June, 1979 to January, 1983 using the Chemical Engineering Plant Cost Index.
5. Table 6.1-4 unit costs were utilized to update O&M costs.
6. The algorithm cost basis was used for working capital, overhead, and capital charge estimation.

- Coal: Eastern bituminous, 11,026 Btu/lb, 3.6 percent sulfur, 10.3 percent ash.
- Limestone handling: truck delivery, 7 days storage
- Solid waste disopsal: hauled by truck to offsite storage
- Ca/S ratio: 2.5
- December 1980 cost basis
- Gulf coast location
- Boiler efficiency of 85 percent

The major adjustments made to the FW estimate to achieve compatibility with the FBC algorithm projections are summarized in Table C-2. These adjustments translated to a total capital cost of $31,110,000 and a total annual cost of $12,250,000. It should be noted that the extensive list of adjustments listed in Table C-2 is due primarily to scope and plant boundary differences between the FW and ITAR estimates, particularly as they effect ancillary equipment. After adjusting costs to a common basis with respect to time of construction, location, and size, the direct capital cost difference for major equipment items (including the boiler fans, ducts, mechanical collector, baghouse, stack, feeders, crushers, limestone handling and storage system, spent solids/ash handling and storage system, and instrumentation) was less than eight percent.

C.3 WESTINGHOUSE ESTIMATE

Westinghouse has estimated FBC capital and operating costs for new industrial boilers over a range of boiler sizes, coal types, and final emission levels.[3] For comparison purposes, the Westinghouse case

TABLE C-2. MAJOR ADJUSTMENTS TO THE FOSTER WHEELER COST BASIS

1. A load factor of 0.6 (as opposed to 0.9) was used to determine annual costs; a labor factor of 0.75 was applied.

2. Guard labor was subtracted from operating labor requirements.

3. Capital costs were adjusted from a Gulf coast to Midwest basis using a factor of 1.028.[a]

4. Capital costs were updated from December 1980 to January 1983, using the Chemical Engineering Plant Cost Index.

5. Table 6.1-4 unit costs were utilized to update O&M costs.

6. The algorithm cost basis was used for land, working capital, overhead, and capital charge estimation.

7. Substituted ITAR coal handling system costs for FW costs since FW design basis included live storage, dead storage, and reclaim equipment. This design basis was significantly more elaborate than the ITAR basis.

8. Substituted ITAR makeup water treatment and chemical feed system costs for FW costs since FW estimate assumed 50 percent makeup water requirement while the ITAR design basis assumed a 20 percent requirement. More importantly, the FW design basis includes a wastewater treatment system which process the following streams:

   - Rainwater runof from paved areas and coal pile.
   - Boiler blowdown.
   - Demineralized regeneration systems.
   - Sanitary waste.

   This equipment is not included within the ITAR plant boundaries.

9. Substituted ITAR cost estimates for the deaeration, boiler feed pumps, and condensate system in place of the FW estimate due to significant differences in design basis.

10. Substituted ITAR cost estimates for buildings and support facilities in place of the FW estimates due to significant differences in scope.

11. Added a 20 percent allowance for contingencies to the FW capital cost estimate.

corresponding to 200 million Btu/hr boiler achieving 80 percent SO_2 removal on a high sulfur Eastern coal has been selected. Three boiler modules are specified for this case. Costs for the boiler and solids (coal, limestone, and bed drain) handling are based on Westinghouse cost files; costs for PM control equipment come from literature sources; costs for boiler auxilliaries are based on PEDCo estimates. Important design factors in the Westinghouse estimate include:

- Air emission standards:
    - 1.2 lb $SO_2/10^6$ Btu
    - 0.5 lb $NO_x/10^6$ Btu
    - 0.03 lb $PM/10^6$ Btu
- Steam conditions: 110 psig at 750°F
- Coal: Eastern bituminous, 11,800 Btu/lb, 3.5 percent sulfur, 10.6 percent ash.
- Coal and limestone handling: Not specified but assumed to be consistent with FBC-ITAR.
- Ca/S ratio: 2.09
- June 1978 cost basis
- Mid-west location
- Boiler efficiency of 84.3 percent

The Westinghouse cost basis is consistent, for the most part, with the ITAR basis. Five modifications to the W estimate were required to achieve consistency with the FBC algorithm basis, as shown in Table C-3. After

TABLE C-3. MAJOR ADJUSTMENTS TO THE WESTINGHOUSE ESTIMATE

1. A labor factor of 0.75 was applied to operating, supervisory, and maintenance labor costs.
2. An allowance for performance tests (1 percent of total direct costs) was added.
3. Capital costs were updated from June 1978 to January 1983 using the Chemical Engineering Plant Cost Index.
4. O&M costs were updated using the unit costs of Table 6.1-4.
5. The algorithm cost basis was used to estimate working capital, overhead, and capital charges.

making these adjustments, the Westinghouse capital cost estimate amounts to $16,760,000; the total annual estimate is $7,579,000/year.

C.4 POPE, EVANS AND ROBBINS ESTIMATE

PER estimated the costs for new FBC boilers at six locations in the Northeast and Midwest to replace existing oil/gas fired boilers.[4] Although costs for cogeneration of steam and electric power were also calculated, only steam generation costs are used for comparison purposes. Heat inputs to the plants were not specified but were estimated from the steam rate, steam conditions, and an assumed boiler efficiency of 85 percent. The case selected for comparison generates 280,000 lb/hr steam at 325 psig (saturated) for an equivalent heat input of 325 million Btu/hr. A Midwest location is assumed. Other particulars of the design basis include:

- Air emission standards: Not specified but assumed to be NSPS for boilers capacities greater than 250 million Btu/hr.

- Three boilers are specified, each rated at 50 percent of total capacity.

- 1979 cost basis.

Insufficient information was provided in the PER estimate description to make adjustments for annual costs. Major adjustments to the PER capital costs to achieve consistency with the FBC algorithm cost basis are summarized in Table C-4. These adjustments resulted in a total capital cost estimate of $31,365,000.

C.5 JOHNSTON BOILER COSTS

JB provided actual installed costs for a 50 million Btu/hr FBC unit operating on Ohio 3.2 percent sulfur coal and controlling SO_2 emissions to

TABLE C-4. MAJOR ADJUSTMENTS TO THE POPE, EVANS AND ROBBINS ESTIMATE

1. Capital cost basis was adjusted to two boilers instead of three as specified.
2. Capital costs were updated from mid-1979 to January 1983 using the Chemical Engineering Plant Cost Index.

2.6 lb $SO_2/10^6$ Btu with limestone.[5] The boiler delivers 50,000 lb steam/hr at 120 psig.

JB provided installed equipment costs for the FBC boiler, baghouse, instrumentation, and auxiliaries. These costs were within 13 percent of the algorithm estimate for a similar boiler. A total capital cost estimate of $4,867,000 was developed by adding algorithm estimates for indirect costs, contingencies, land, and working capital to the JB installed equipment costs. No other adjustments are necessary as the JB costs conform to a December 1982 basis.

Insufficient information was provided with the JB cost description to make adjustments for annual costs.

APPENDIX C REFERENCES

1. Myrick, D. T. (Combustion Engineering, Inc.) DOE Cost Comparison Study: Industrial Fluidized Bed Combustion VS. (Conventional Coal Technology. (Prepared for U. S. Department of Energy.) FE-2473-T7. January 1980.

2. Foster Wheeler Development Corporation. Industrial Steam Supply System Characteristics Program, Phase 1, Conventional Boilers and Atmospheric-Fluidized-Bed Combustor. (Prepared for Oak Ridge National Laboratory, U. S. Department of Energy). ORNL/Sub-80/13847/1. August 1981.

3. Ahmed, M. M., D. L. Keairns, and R. A. Newby (Westinghouse Research and Development Center). Effect of Emission Control Requirements on Fluidized-Bed Boilers for Industrial Applicators: Preliminary Technical/Economic Assessment. (Prepared for U. S. Environmental Protection Agency.) EPA-600/7-81-149. September 1981.

4. Mesko, J. E. (Pope, Evans and Robbins Inc.). Economic Evaluation of Fluidized Bed Coal Burning Facilities for Industrial Steam Generation. The Proceedings of the Sixth International Conference on Fluidized Bed Combustion, Volume II. Atlanta, Georgia. August 1980.

5. Letter from Virr, M. J., Johnston Boiler Company, to Aul, E. F., Radian Corporation. November 18, 1983. FBC boiler cost study.

APPENDIX D
BASES FOR COST ESTIMATES

D.1 COSTING METHODOLOGY

Costs for model boilers have been developed on the basis of construction and operation in the Midwest region of the U.S. Although the absolute costs for model boilers and various SO_2 control alternatives will vary from region to region, the cost differentials between alternatives are not expected to differ significantly on a regional basis. For the purposes of this report, costs have been developed for the Midwest region only.

All costs in this report are presented on a January 1983 basis, except where noted.

The costs of each model boiler can be broken down into three major cost categories:

- Capital Costs (total capital investment required to construct and make operational a boiler and control system),
- Operation and Maintenance (O&M) costs (total annual cost necessary to operate and maintain a boiler and control system), and
- Annualized Costs (total O&M costs plus capital-related charges).

Each of these cost categories can be further subdivided into individual cost components.

Capital Costs

Table D-1 presents the individual capital cost components and the general methodology used for calculating total capital costs. The plant boundaries include inlets to coal and sorbent storage, boiler feedwater inlet to the economizer, steam outlets from the steam generator, on-site

TABLE D-1. CAPITAL COST COMPONENTS[a]

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(1) Direct Costs

Equipment
+ Installation
= Total Direct Costs

(2) Indirect Costs

Engineering - 10% of direct costs for boilers and PM controls[b]

For FGD systems on boilers $<200 \times 10^6$ Btu/hr, FGD engineering costs are 10% of FGD direct costs for an FGD system that is applied to a 200×10^6 Btu/hr boiler.

For FGD systems on boilers $\geq 200 \times 10^6$ Btu/hr, FGD engineering costs are 10% of specific FGD system's direct costs.[c]

+ Construction and Field Expenses (10% of direct costs)[b]
+ Construction Fees (10% of direct costs)[b]
+ Start Up Costs (2% of direct costs)[b]
+ Performance Costs (1% of direct costs)[c]
= Total Indirect Costs

(3) Contingencies[b] = 20% of (Total Indirect + Total Direct Costs)

(4) Total Turnkey Cost = Total Indirect Cost + Total Direct Cost + Contingencies

(5) Working Capital[1] = 25% of Total Direct Operating Costs[d]

(6) Land[e]

(7) Total Capital Cost = Total Turnkey + Working Capital + Land

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[a] Boiler and each control system costed separately; factors apply to cost of boiler or control system considered; i.e., the engineering cost for the PM control system is 10% of the direct cost of the PM control system.

[b] Reference 1.

[c] Reference 2.

[d] This equation is used for control device working capital calculations. For boilers, fuel supplies are included so a different equation is used (see Table D-2).

[e] Land costs are assumed to apply to boilers only.

spent solids storage outlets, and the stack outlet. The costs for the steam and condensate return lines from the process area are not included. Battery limits of the emissions control systems include the control devices themselves, raw material handling, temporary waste storage, and any additional ducting required.

Direct capital costs consist of the basic and auxiliary equipment costs in addition to the labor and material required to install the equipment. Indirect costs are those costs not attributable to specific equipment items. Other capital cost components are contingencies, the cost of land, and working capital.

Contingencies are included in capital costs to compensate for unpredicted events and other unforeseen expenses. Costs for land are included in boiler capital costs but not in control system costs. All boilers except pulverized coal boilers are assumed to have land costs of $2,800. Pulverized coal boilers are assumed to have land costs of $5,700.[1]

The computation of working capital in this analysis also differs slightly between boilers and control equipment. The equations shown in Table D-2 are used to calculate the cost for working capital. These equations are based on three months of direct annual non-fuel operating costs and one month of fuel costs.

Operation and Maintenance (O&M) Costs

Table D-3 lists the individual O&M cost components and the general methodologies used in calculating total O&M costs. Direct O&M costs include operating and maintenance labor, fuel, utilities, spare parts, supplies, waste disposal and chemicals. Indirect operating costs include payroll and plant overhead and are calculated based on a percentage of some key O&M cost components (e.g. direct labor, supervisory labor, maintenance labor and spare parts).

TABLE D-2. WORKING CAPITAL CALCULATIONS FOR BOILERS AND CONTROL DEVICES

Working Capital (WC)

Boilers - Assume three months of direct annual non-fuel operating costs and one month of fuel costs

$$WC^{a} = 0.25\ (\text{Direct annual non-fuel operating costs}) + 0.083\ (\text{Fuel costs})$$

Control Equipment - Assume three months of direct annual operating costs

$$WC^{b} = 0.25\ (\text{Direct annual operating costs})$$

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[a] Reference 3.

[b] Reference 1.

TABLE D-3. OPERATING AND MAINTENANCE COST COMPONENTS[a]

(1) Direct Operating Costs

Direct Labor
+ Supervision
+ Maintenance Labor, Spare Parts and Supplies
+ Electricity
+ Water
+ Steam
+ Waste Disposal
  Solids (Fly ash and bottom ash)
  Sludge
  Liquid
+ Chemicals

Total Non-Fuel O&M
+ Fuel

= Total Direct Operating Costs

(2) Indirect Operating Costs (Overhead)[b]

Payroll (30% Direct Labor)
+ Plant (26% of Direct Labor + Supervision + Maintenance Costs + Spare Parts)

(3) Total Annual Operating and Maintenance Costs = Total Direct + Total Indirect Costs

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[a] Boilers and each control systems are costed separately; factors apply to boiler or control system being considered, (i.e., payroll overhead for FGD system is 30% of the direct labor requirement for the FGD system).

[b] Factors recommended in Reference 4.

The key factors used in calculating annual O&M costs are the system capacity utilization, utility unit costs (steam, electricity, water), and unit costs for raw materials, waste disposal, and labor. Capacity utilization is defined as the actual annual fuel consumption as a percentage of the potential annual fuel consumption at maximum firing rate. Table D-4 presents the utility and unit costs used in calculating annual O&M costs for the boilers and control equipment.

The largest O&M cost for boilers is fuel. Table 6.1-3 presents the specifications and costs for the fuels used in this analysis. To maintain consistency with the Industrial Fuel Choice Analysis Model (IFCAM), which is used to project the national impacts of alternative SO_2 standards, the values in Table 6.1-3 are projections for 1990 delivered fuel prices expressed in January 1983 dollars.[7,8] These projections ignore the effects of inflation but assume that fuel prices will escalate in real terms. In addition, the fuel prices have been "levelized" over the life of the boiler (i.e., an equivalent constant price has been calculated after allowing for escalation and the time value of money).

Annualized Costs

Total annualized costs are the sum of the annual O&M costs and the annualized capital charges. The annualized capital charges include the payoff of the capital investment (capital recovery), interest on working capital, general and administrative costs, taxes, and insurance.

Table D-5 presents the methods used in this report to calculate the individual annualized capital charges components. The capital recovery cost is determined by multiplying the capital recovery factor, which is based on the real interest rate and the equipment life, by the total turnkey costs (see Table D-1). For this analysis, a 10 percent real interest rate and a 15 year equipment life are assumed for the boilers and control equipment. This translates into a capital recovery factor of 13.15 percent. The real

TABLE D-4. UNIT COSTS USED IN MODEL BOILER CALCULATIONS[a]

Utilities	
Electricity	0.0503/kwh[b]
Water	0.0396/m^3 (\$0.15/10^3 gal)[c]
Steam	\$3.5/10^3 lb[d]
Raw Materials	
Na_2CO_3	\$0.169/kg (\$153/ton)[c,e]
Lime	\$0.098/kg (\$89/ton)[c,e]
Limestone	\$0.013/kg (\$8.5/ton)[c]
Labor	
Direct Labor	\$11.75/man-hour[f,g]
Supervision	\$15.28/man-hour[h]
Maintenance Labor	\$14.34/man-hour[i]
Waste Disposal	
Solids (Ash, Spray Dried Solids)	\$0.198/kg (\$18/ton)[j,h]
Sludge	\$0.0198/kg (\$18/ton)[j]

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[a] All costs in January 1983 \$.

[b] Monthly Energy Review, April 1983.

[c] TVA, Technical Review of Dry FGD Systems and Economic Evaluation of Spray Dryer FGD Systems, February 1982.

[d] EPRI, Technical Assessment Guide, May 1982.

[e] Updated using ratio of commodity chemical price for January, 1983 to June, 1982 as given in the Chemical Marketing Reporter.

[f] Monthly Labor Review April, 1982.

[g] Average of wate rates for Chemical and Allied Products and Petroleum and Coal Products categories.

[h] Estimated at 30 percent over direct labor rate.

[i] Estimated at 22 percent over direct labor rate.

[j] Average of waste disposal rates from EPA, Economics of Ash at Coal Fired Power Plants, Oct. 1981, and EEA, Estimated Landfill Credit for Non-Fossil Fueled Boilers, October, 1980.

TABLE D-5. ANNUALIZED COST COMPONENTS

(1) Total Annualized Cost = Annual Operating Costs + Capital Charges

(2) Capital Charges = Capital recovery + interest on working capital + miscellaneous (G&A, taxes and insurance)

(3) Calculation of Capital Charges Components

A. Capital Recovery = Capital Recovery Factor (CRF) x Total Turnkey Cost

$$CRF = \frac{i\,(1 + i)^n}{(1 + i)^n - 1}$$

i = interest rate

n = number of years of useful life of boiler or control system

Item	n	i	CRF
Boiler, control systems	15	10	0.1315

B. Interest on Working Capital = 10% of working capital

C. G&A, taxes and insurance = 4% of total turnkey cost interest rate of 10 percent was selected as a typical constant dollar rate of return on investment to provide a basis for calculation of capital recovery charges. This interest rate is the "real" interest rate above and beyond inflation.

Table D-5 also presents the methods to calculate the other annualized capital charges components. Interest on working capital is based on a 10 percent interest rate. The remaining components (general and administrative costs, taxes, and insurance) are estimated as 4 percent of total turnkey costs.

D.2 BOILER AND CONTROL COST PARAMETERS

Capital and annualized costs for model boilers and PM, NO_x, and SO_2 control techniques are estimated in this report by the use of cost "algorithms". Each algorithm is an algebraic function which projects capital and annual costs for a particular system based on key process parameters (e.g., heat input to boiler, SO_2 removal efficiency, capacity utilization factor, flue gas flow rate). The algorithms have been computerized to allow rapid and accurate cost calculations over a wide range of boiler/control system size ranges and operating conditions. Summary information describing the boiler and emission control costing algorithms used in this report is presented in Table D-6. A complete listing of the algorithms is provided in Appendix A and Reference 21. The specific equipment lists and assumptions used to develop the various algorithms are discussed in the following sections.

Boiler Costs

This section presents the specific cost assumptions and methodologies that were used to calculate the industrial boiler costs presented in Section 6.0. References 9 and 10 detail the specific equipment lists and assumptions used to develop the boiler algorithms presented in Appendix A and Reference 21 .

TABLE D-6. SUMMARY OF BOILER AND EMISSIONS CONTROL COSTING ALGORITHMS

Abbreviation	Algorithm Type	Boiler Size Applicability MW (10^6 Btu/hr)
UNDR	Boiler, underfeed stoker, watertube, package	$\leq$22 ($\leq$75)
SPRD	Boiler, spreader stoker, watertube, field-erected	18 - 58 (60 - 200)
PLVR	Boiler, pulverized coal, watertube, field-erected	$\geq$58 $\leq$200)
FBC	Boiler, fluidized bed, watertube, shop fabricated	8.8 - 117.2 (30 - 400)
FF	Fabric filter applied to coal-fired boiler	8.8 - 204 (30 - 700)
DS	Lime spray drying (dry scrubbing) FGD system	All sizes
LEA	Low excess air operation for NO_x control	All sizes
SCA	Staged combustion air applied to coal-fired boilers	$\geq$44 ($\geq$150)

As mentioned previously, the capacity utilization factor and labor factor are used to adjust O&M costs for boiler operation at less than full capacity. The factors used in this report are summarized in Table D-7. These factors are considered representative of industrial boiler operation, and are supported by information in References 3 and 11. The capacity utilization and labor factors shown in Table D-7 are also used to adjust O&M costs for PM, NO_x, and SO_2 controls.

The boiler specifications presented in Table D-8 have been used to calculate the conventional boiler capital costs presented in this report. It is assumed that all boilers are operating under low excess air firing conditions. The flue gas flow rates presented were calculated from applicable algorithms.

2.3.2 Particulate Matter (PM) Control Costs

The algorithms used to calculate capital and operating costs for PM control devices are presented in Reference 21. The cost algorithms for reverse-air fabric filters were developed by PEDCo, Inc. Detailed documentation of the cost bases for these controls can be found in PEDCo's final report.[12,13] Table D-9 lists the general specifications for the PM control devices investigated. These specifications are typical for industrial boiler control devices currently in use.

NO_x Control Costs

The algorithms used to calculate capital and operating costs for NO_x control devices are presented in Reference 21. The cost algorithms for low excess air (LEA) operation, and staged combustion (SCA) were developed by Radian based on costs presented in the Individual Technology Assessment Report (ITAR) for NO_x Combustion Modification.[14] Table D-10 presents the general specifications for LEA and SCA.

TABLE D-7. CAPACITY UTILIZATION AND LABOR FACTORS USED FOR MODEL BOILER COST CALCULATIONS[a]

Boiler Type	Capacity Utilization Factor (CF)	Labor Factor (LF)
Coal-fired (Underfeed, spreader stoker, pulverized feed)	0.60	0.75

Labor Factor Equations

CF	LF
>0.7	1
0.5 - 0.7	0.5 + 2.5 (CF - 0.5)
<0.5	0.5

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[a] References 3 and 11.

TABLE D-8. SPECIFICATIONS FOR CONVENTIONAL COAL-FIRED BOILERS

Thermal input, MW (10^6 Btu/hr)	14.5 (50)	29.0 (100)	44.0 (150)	73.0 (250)	117.2 (400)
Fuel firing method	Underfeed stoker	Spreader stoker	Spreader stoker	pulverized coal	Pulverized coal
Fuel analysis					
Percent sulfur	3.23	3.23	3.23	0.42	0.42
Percent ash	12.0	12.0	12.0	6.9	6.9
Heating value, kJ/kg (Btu/lb)	27,200 (11,700)	27,200 (11,700)	27,200 (11,700)	20,500 (8,825)	20,500 (8,825)
Excess air, percent	35	35	35	35	35
Flue gas flow rate, m^3/s (acfm)	8.70 (18,400)	17.4 (36,800)	26.0 (55,100)	43.9 (93,000)	67.0 (142,000)
Load factor, percent	60	60	60	60	60
Efficiency, percent	79.0	80.0	80.9	82.0	83.1
Steam production, kg/hr (lb/hr)	17,600 (38,800)	32,000(70,400)	48,500 (106,900)	78,400 (173,000)	127,010 (280,000)

[a] Conditions correspond to low excess air operation.

TABLE D-9. GENERAL DESIGN SPECIFICATIONS FOR PM CONTROL SYSTEMS

Control Device	Item	Specification
Fabric Filter (FF)	Material of construction	Carbon steel (insulated)
	Cleaning method	Reverse-air (multi-compartment)
	Air to cloth ratio	2 acfm/ft^2
	Bag material	Teflon-coated fiberglass
	Bag life	2 years
	Pressure drop[a]	6 in. H_2O

[a] Pressure drop refers to gas side pressure drop across entire control system.

TABLE D-10. NO_x COMBUSTION MODIFICATION EQUIPMENT REQUIREMENTS ON CONVENTIONAL BOILERS

Control Device	Specification
Low Excess Air (LEA)	Oxygen trim system - O_2 analyzer, air flow regulators
	Wind box modifications (may be required for multi-burner boilers)
Staged combustion Air (SCA) Pulverized coal-fired boilers:	Oxygen trim system - O_2 analyzer, air flow regulators
	Airports
	Wind box modifications
	Larger forced draft fan power

SO_2 Control Costs

The cost algorithms used to calculate capital and annual operating costs for flue gas desulfurization units are also presented in Reference 21. The cost basis for the lime spray drying FGD systems is presented in the FGD ITAR. Cost algorithms based on the ITAR cost estimates were developed by Acurex Corporation.[15] The algorithms presented in Reference 21 however, do not represent the costs in the final ITAR or the Acurex report for the spray drying systems. The Acurex algorithms were modified to reflect revised installation factors and revised fabric filter costs for the spray drying systems. These revisions are documented in a several technical memos.[16,17]

Table D-11 presents the general specifications for the lime spray-drying FGD system analyzed in this report. These specifications are typical for lime spray drying systems currently in use.

Liquid and Solid Waste Disposal

The major liquid and solid waste streams from uncontrolled conventional boilers are: water softening sludge, condensate blowdown, bottom ash disposal, and coal pile runoff. Bottom ash collection, handling, and disposal costs have been incorporated into the uncontrolled boiler cost estimates. Bottom ash disposal costs were estimated based on a non-hazardous waste classification under RCRA regulations. If industrial boiler wastes are classified as hazardous in the future, the disposal costs and overall boiler control costs (for coal-fired boilers) would increase significantly.

Disposal of fly ash (from PM control devices), spray dryer solids (from the dry SO_2 scrubbing process), and spent solids (from FBC boilers) has also been estimated on the basis of a non-hazardous waste classification.

TABLE D-11. GENERAL DESIGN SPECIFICATIONS FOR THE LIME SPRAY DRYING FGD SYSTEM

Control Device	Item	Specification
Dry scrubbing (spray drying, SO_2 and PM removal) (DS)	Material of construction	Carbon steel spray dryer and fabric filter (insulated)
	Reagent	Lime; with solids recycle at 2 kg recycle solids/kg fresh lime feed
	Fabric filter	Pulse jet; air-to-cloth ratio of 4 acfm/ft^2
	Pressure drop[a]	6 in. H_2O
	L/G	0.3 gal/acf
	Solids disposal	Trucked to off-site landfill

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[a] All pressure drops refer to gas side pressure drop across entire control system.

Costs for treating the other three waste streams were not quantitatively evaluated in this study. The costs associated with waste stream disposal are highly site-specific and are influenced by the following parameters:

- Water softening sludge rate and composition: raw water quality, steam quality, and water makeup rate.
- Condensate blowdown rate and composition: effluent discharge quality requirements, raw water quality, and condensate blowdown quantity.
- Coal pile runoff rate and composition: coal quality, meterological conditions, and effluent discharge quality requirements.

However, these costs would be associated with the boiler itself and would not affect the analysis of incremental costs for air pollution control systems.

APPENDIX D REFERENCES

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13. Bowen, M.L., (Radian Corporation.) Costs of Mechanical Collectors Applied to Fossil Fuel Fired Industrial Boilers. June 2, 1982. 12 p.

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APPENDIX E

AUXILIARY LISTINGS OF AFBC MANUFACTURERS AND UNITS

As a supplement to the information presented in Section 3.0, this appendix contains summary lists of foreign AFBC manufacturers, existing and planned foreign coal-fired AFBC units, and existing and planned multi-fuel and alternative fuel AFBC units.

TABLE E-1. FOREIGN AFBC MANUFACTURERS[35]

	AFBC Boiler Technology				Boiler Capabilities Commercially Available				Number of Units Installed	
Company Address	Built Under License	Licensing Company	Watertube or Firetube Boiler	Types of FBC Systems Offered	Steam Capacity, 1000 lb/hr	Pressure, psig	Temperature, °F	Fuel(s)	USA	Total
A. Ahlstrom Oy P. O. Box 329 SF-00101 Helsinki 10, Finland	No	-	Wt	Fx, Fcb	20-400	140-2500	350-1000	-[1]	0	9
Ansaldo SpA Viale Sarca, 336 Milano 20126, Italy	No	-	Wt	Fx	Up to 400	NAv	Up to 1000	Coal, Woodwaste	1	0
Babcock Hitachi KK 6-2, 2-Chome, Ota-machi Chiyodo-Ku, Tokyo 100, Japan	No	-	Wt	Fx	22-1100	100-2400	Up to 1050	-[1]	2	2
Combustion Systems Ltd. BP Research Centre Sunbury-on-Thames, Middlesex England TW16 7LN	No	-	Wt, Ft[2]	Fx[2]	25-500[2]	1000-2400[2]	Up to 1005[2]	-[1]	0	0
Danks of Netherton, Ltd. Halesowen Rd, Netherton Dudley, West Midlands England DY2 9PG	Yes	Combustion Systems Ltd.	Wt, Ft	Fx	15-70	100-900	Up to 900	-[1]	0	4
Deborah Fluidized Combustion, Ltd. 6 Davy Dr. NW Industrial Estate Peterlee, Durham, England	No	-	Wt	Fcb	1-50	100-900	-[3]	-[1]	1	12
Deutsche Babcock Werke AG Duisburger Strasse 375 Oberhausen D-4200, W. Germany	No	-	Wt	Fx, Pcb	20-700	145-2600	360-1100	-[1]	0	14
Fluidised Combustion Contractors Ltd. 11 The Boulevard Crawley, Sussex England RH10 1UX	Yes	Solids Circulation Systems, Inc.[4]	Wt, Ft	Fx, Pcb, Fcb	-[5]	-[5]	-[5]	-[1]	1	2
Foster Wheeler Power Products Ltd. Greater London House Hampstead Rd., London England NW1 7QN	Yes	-[6]	Wt	Fx, Fcb	30-600	200-2000	200-1000	-[1]	0	2
Generator Industrie AB P. O. Box 95 S-433 22 Partille, Sweden	Yes	Fluidized Combustion Co.	Wt	Fx, Fcb	17-170	150-1000	Up to 800	Coal, Woodwaste, Biomass	0	8

TABLE E-1. FOREIGN AFBC MANUFACTURERS[35] (Continued)

Company Address	AFBC Boiler Technology		Watertube or Firetube Boiler	Types of FBC Systems Offered	Boiler Capabilities Commercially Available				Number of Units Installed	
	Built Under License	Licensing Company			Steam Capacity, 1000 lb/hr	Pressure, psig	Temperature, °F	Fuel(s)	USA	Total
E. Green & Son Ltd. Wakefield England WF1 5PF	No	-	Wt	Fx	20-80	150-900	Up to 900	-[1]	1	1
Ishikawajima-Harima Heavy Industries Co., Ltd. 30-13 5-Chome, Toyo Koto-ku, Tokyo 135, Japan	Yes	Fluidized Combustion Co.	Wt, Ft	Fx	-[5]	-[5]	-[5]	-[1]	0	2
NE Boilers Ltd. NE House, Fengate Peterborough, Cambs. England PE1 5BQ	Yes	Combustion Systems, Ltd.	Wt	Fx	20-100	Up to 2500	Up to 900	-[1]	0	1
NEI Cochran Ltd. Newbie Works Annan, Dumfriesshire Scotland DG12 5QU	No	-	Ft	Fx	2-36	100-250	Sat	-[1]	0	6
Tampella Ltd., Boiler Div. P. O. Box 626 SF-33101 Tampera 10, Finland	No	-	Wt	Fx	13-225	400-1800	Up to 1000	-[1]	0	6
Wallsend Slipway Engineers Ltd. Point Pleasant Wallsend, Tyne & Wear England NE28 6QN	No	-	Ft	Fx	5.3-59.8	150-250	Sat	-[1]	0	0

Footnotes:

1. Designed to burn the following fuels separately or in combination: coal, woodwaste, biomass, liquid wastes or sludges, coal-washing wastes.
2. Range of equipment specifications offered.
3. Temperature depends on customer requirements.
4. Fluidized Combustion Contracts Ltd. offers fluidized bed combustion systems of its own design as well.
5. Designed to meet customer requirements.
6. Foster Wheeler Power Products Ltd. licenses the fluidized-bed technology for some of the equipment it offers from Fluidized Combustion Co., a joint venture of Foster Wheeler Development Corp. and Pope Combustion Systems Inc., and from Battelle Memorial Institute

Abbreviations:

Fcb--Full circulating bed

Ft--Firetube boiler

Fx--Fixed (bubbling) bed

NAv--Not available

Pcb--Partial circulating bed

Sat--Saturation temperature

Wt--Watertube boiler

TABLE E-2. EXISTING AND PLANNED FOREIGN COAL-FIRED AFBC UNITS[35]

Plant Owner	Location	Steam Capacity, 1000 lb/hr	Steam Pressure psig	Steam Temperature °F	Design Fuel(s)	Manufacturer	Type of Project	Type of Financing	Commercial Service Date
Atlas Consol Mining & Dev. Corp.	Cebu, Philippines	352[1]	914	905	L	DBW[2]	Com	P	1982
Elektrizitatswerk Wesertral GmbH	Hamlin, W. Ger.	309	1741	986	C	DBW[2]	Com	P/G	1983
Dibso Power Plant	China	286	588	840	C	-	Com	G	4/80
Saarbergwerke AG	Volklingen, W. Ger.	273[1]	...[3]	...[3]	C	DBW[2]	Com	P/G	1982
National Coal Board	Grimethorpe, Eng.	176	435	824	C	DBW[2]	D	G	1980
ENEL[4]	Porto Vesma, Italy	175	840	890	C	ANS	D	P/G	1984
Shell Nederland Raffinaderji BV	Pirnus, Holland	110	1174	923	C	FWC	Com	P	7/82
Jiangapen	Guangdong, China	110	605	794	C	-	Com	G	1981
Ruhrkohl AG	Dusseldorf, W. Ger.	109	247	752	C	DBW[2]	D	P/G	1979
British Steel Corp.	Sheffield, Eng.	80	650	820	C	MEB	Com	P/G	7/81
Yuiyang	Hunan, China	77	650	794	C	-	Com	P/G	1981
Mitsui Toatsu Chemicals, Inc.	Sunagawa, Japan	69	356	536	C	IHI	Com	P	4/82
City of Vastervik	Vastervik, Sweden	68	175	375	C	GEN	Com	P	12/83
Babcock Power Ltd.	Renfrew, Scotland	60	400	518	C	FCL	D	P	5/75
Mitsui	Toatsu Chan, Japan	55	356	482	C	-	Com	P/G	NAv

TABLE E-2. EXISTING AND PLANNED FOREIGN COAL-FIRED AFBC UNITS[35] (Continued)

Plant Owner	Location	Steam Capacity, 1000 lb/hr	Steam Pressure psig	Steam Temperature °F	Design Fuel(s)	Manufacturer	Type of Project	Type of Financing	Commercial Service Date
Babcock Hitachi KK	Wakamatsu, Japan	44	853	1000	C	HIT	D	P/G	4/81
Chalmers University	Gothenburg, Sweden	44	580	800	C	GEN	Com	P/G	3/82
Canadian Dept. of Defense	Summersido, PWI, Can.	40[1]	160	Sat	C	FWC	D	G	12/82
Mooming Petroleum	China	32	180	482	C	-	Com	G	12/65
Tsinghum University	Beiding, China	30	336	734	C	-	Com		6/64
Chemical Plant Cogen	Trichy, India	26	200	480	C	BHEL	Com	P/C	10/81
Undisclosed	Undisclosed	24	384	Sat	C	HIT	Com	P	1984
Danks of Netherton Ltd.	Dudley, Eng.	20	400	Sat	C	DNL	D	P/G	NAv
Hastra	Luneburg, W. Ger.	19[1]	885	923	C	DBW	Com	P	1983
Saarbergwerke AG[5]	Volklingen, W. Ger.	17[3]	...[3]	...[3]	C	DBW[2]	D	P	1980
Danks Engineering Ltd.	Oldbury, Eng.	16	150	Sat	C	DNL	Com	P	5/81
Smith's Brewery Ltd.	Tadeaster, Eng.	15	150	Sat	C	NEI	Com	P	1981
Sulzer Brothers Ltd.	Winterthur, Switzerland	12	435	572	C	SUL	D	P/G	9/79
E. Green & Son Ltd.	Wakefield, Eng.	10	180	Sat	C	GRE	D	P	6/82
Undisclosed	Undisclosed	10	150	Sat	C	DNL	Com	P	5/82
National Coal Board	Selby, Eng.	4[6]	50	Sat	C	NEI	Com	P/G	1981
North York County Council	Knaresborough, Eng.	4[7]	NAv	NAv	C	DFC	NAv	NAv	1982

FOOTNOTES, ABBREVIATIONS, AND MANUFACTURERS FOR TABLE E-2

Footnotes:

1. Two units installed.
2. In conjunction with Vereinigte Kesselwerke AG.
3. Rating is in million Btu/hr; hot combustion gas exiting fluidized-bed combustor flows to a conventional fired boiler.
4. Steam at 80 percent quality.
5. Prototype power station.
6. Four units installed.
7. Rating is in million Btu/hr; unit is a fluidized-bed hot-water boiler. Operating pressure and temperature are for hot water.

Abbreviations:

C--Coal

Com--Commercial contract

D--Demonstration project

G--Government financing

NAv--Not available

P--Private financing

P/G--Private/government financing

Sat--Saturated

FOOTNOTES, ABBREVIATIONS, AND MANUFACTURERS FOR TABLE E-2 (Continued)

Manufacturers:

ANS--Ansaldo SpA

BHEL--Bharat Heavy Electricals Ltd.

DBW--Deutsche Babcock Werke AG

DFC--Deborah Fluidised Combustion Ltd.

DNL--Danks of Netherton Ltd.

FCL--Fluidized Combustion Contractors Ltd.

FWC--Foster Wheeler Boiler Corporation

GEN--Generator Industrie AB

GRE--E. Green & Sons Ltd.

HIT--Babcock Hitachi KK

IHI--Ishikawajima-Harima Heavy Industries Co.

MEB--M E Boilers Ltd.

NEI--NEI Cochran Ltd.

SUL--Sulzer Brothers Ltd.

TABLE E-3. EXISTING AND PLANNED MULTI-FUEL AND ALTERNATE FUEL AFBC UNITS[35]

Plant Owner	Location	Steam Capacity, 1000 lb/hr	Steam Pressure psig	Steam Temperature °F	Design Fuel(s)	Manufacturer	Type of Project	Type of Financing	Commercial Service Date
Ashland Petroleum Company	Catlettsburg, Ky.	325[1]	450	700	CO,Ng	FWC	Com	P	2/83
A Ahlstrom Oy[2]	Kauttus, Finland	200	1200	930	Pt,C	AHL	Com	P	4/81
Kemira Oy	Oulu, Finland	155	1275	960	Pt,C	AHL	Com	P	1/83
Zellstoff-und Papierfabrik AG[2]	Frantschach, Austria	154	1215	970	W,Bc	AHL	Com	P	11/83
Northern States Power Co.	LaCrosse, Wis.	150	450	750	W	EPI	Com	P	12/81
Hylte Bruks AB	Hyltebruk, Sweden	143	925	840	Pt,W,C	AHL	Com	P	8/82
Dortmund Colliery	Dortmund, W. Ger.	73	485	797	Cww	DBW	Com	P/G	2/82
Flingorn Power Station	Dusseldorf, W. Ger.	110	250	750	Bc	DBW		P/G	1980
Undisclosed	Undisclosed	93	327	Sat	PrW	IHI	Com	P	4/83
Hyvinkaan Lampovoima Oy	Hyvinkaa, Finland	85[3]	130[3]	355	Pt,C,W	AHL	Com	P	9/81
Kirby Lumber Co.	Silsbee, Texas	70	350	Sat	W	EPI	Com	P	12/80
City of Gallivare	Gallivare, Sweden	68	232	356	Pt	TAM	Com	P	9/83
American Can Co.	Bellamy, Ala.	55	150	Sat	W	YSI	Com	P	4/80
State of California	Sacramento, Calif.	45	275	Sat	W	EPI	Com	G	10/82
DeArmond Stud Mill	Coeur d'Alene, Idaho	40	150	Sat	W	EPI	Com	P	6/78
E. Stroudsburg State Coll.	E. Stroudsburg, Pa.	40	150	Sat	Ac	FEC	D	P/G	6/83
Weyerhaeuser Co.	Raymond, Wash.	40	150	Sat	W	EPI	Com	P	11/75
Atlantic Veneer Corp.	Beaufort, N. C.	35	200	Sat	W	YSI	Com	P	5/77
City of Eksjo	Eksjo, Sweden	34	115	340	W	GEN	Com	P	2/81
Idaho Forest Industries	Coeur d'Alene, Idaho	30	150	Sat	W	EPI	Com	P	9/73
Sumter Plywood Corp.	Livingston, Ala.	27	180	Sat	W	EPI	Com	P	12/77
Northwestern Mississippi Jr. College	Senatobia, Miss.	27	150	Sat	W	EPI	Com	P	3/80

TABLE E-3. EXISTING AND PLANNED MULTI-FUEL AND ALTERNATE FUEL AFBC UNITS[35] (Continued)

Plant Owner	Location	Steam Capacity, 1000 lb/hr	Steam Pressure psig	Steam Temperature °F	Design Fuel(s)	Manufacturer	Type of Project	Type of Financing	Commercial Service Date
Boise Cascade Corp.	Emmett, Idaho	26	150	Sat	W	EPI	Com	P	3/77
Boise Cascade Corp.	Moncur, N. C.	26	150	Sat	W	EPI	Com	P	11/77
Webster Lumber Co.	Bangor, Wis.	26	150	Sat	W	EPI	Com	P	3/77
Diamond International Corp.	Redmond, Ore.	25	150	Sat	W	EPI	Com	P	12/80
Atlantic Veneer Corp.	Beaufort, N. C.	24[4]	200	Sat	W	YSI	Com	P	3/81
Shamokin Area Ind. Corp.	Shamokin, Pa.	24	200	Sat	Ac	KEE	D	P	10/81
Kogap Manufacturing Co.	Medford, Ore.	24	180	Sat	W	EPI	Com	P	4/79
Skelleftea Kraft AB	Skelleftea, Sweden	24[3]	130[3]	355[3]	Pt	AHL	Com	P	12/81
Savon Voima Oy	Suonerjoki, Finland	24[3]	130[3]	250[3]	Pt	AHL	Com	P	11/79
Sumitomo Coal Mining Co.	Akabira City, Japan	22	100	Sat	Cww	HIT	Com	P	4/79
Nagel Lumber Co.	Land O'Lakes, Wis.	21	175	Sat	W	YSI	Com	P	8/77
Wade Lumber Co.	Wade, N. C.	21	150	Sat	W	YSI	Com	P	6/79
Chapleau Lumber Co.	Chapleau, Ont., Can.	21	15	Sat	W	YSI	Com	P	2/77
Superwood Corp.	Phillips, Wis.	20	250	Sat	W	EPI	Com	P	7/77
Eastmont Forest Products	Ashland, Mont.	20	150	Sat	W	EPI	Com	P	3/74
Merritt Brothers Lumber Co.	Priest River, Idaho	20	150	Sat	W	EPI	Com	P	1/76
Multnomah Plywood Corp.	St. Helens, Ore.	20	150	Sat	W	EPI	Com	P	9/79
City of Eksjo	Eksjo, Sweden	17	115	340	R	GEN	Com	P/G	12/79
H&B Lumber Co.	Marion, N. C.	14	150	Sat	W	YSI	Com	P	11/75
Undisclosed	Undisclosed	12	150	Sat	NAv	NEI	Com	P	1982
Tenneco Ltd.	Bristol, Eng.	10	250	Sat	Wt	DML	Com	P	8/80
Binghamton Psychiatric Center	Binghamton, N. Y.	10	150	Sat	W	DED	Com	P	11/80

TABLE E-3. EXISTING AND PLANNED MULTI-FUEL ALTERNATIVE FUEL AFBC UNITS[35] (Continued)

Plant Owner	Location	Steam Capacity, 1000 lb/hr	Steam Pressure psig	Steam Temperature °F	Design Fuel(s)	Manufacturer	Type of Project	Type of Financing	Commercial Service Date
Boise Cascade Corp.	Cascade, Idaho	10	150	Sat	W	EPI	Com	P	3/80
Lindsay Olive Growers	Lindsay, Calif.	10	150	Sat	Op	EPI	Com	P	4/76
Rossi Corp.	Higganum, Ct.	10	150	Sat	W	YSI	Com	P	12/79
Kelly Enterprises	Pittsfield, Mass.	10	15	Sat	W	YSI	Com	P	2/75
Walnut Products, Inc.	St. Joseph, Mo.	9	150	Sat	W	YSI	Com	P	10/75
Iowa-Missouri Walnut Co.	St. Joseph Mo.	7	150	Sat	W	YSI	Com	P	10/75
Undisclosed	Haifa Bay, Israel	60	200	Sat	Ch,PrW	EPI	Com	P	1982
Oy Alko Ab	Koskenkorva, Finland	56	585	840	Pt,O	AHL	Com	P	1/83
City of Lisalmi	Lisalmi, Finland	51[3]	232[3]	356[3]	Pt,W	TAM	Com	P	11/83
City of Scandvikan	Scandvikan, Sweden	51[1]	175	375	W,C,Pt	GEN	Com	P	11/83
Conoco, Inc.	Uvalde, Texas	50	2450	665	C,L,Ck	SMC	Com	P	12/81
Campbell Soup Co.	Maxton, N. C.	150[5]	300	Sat	C,PrW[6]	JBC	Com	P	10/82
Stevenson Dyers Ltd.	Ambergate, Eng.	50	250	460	C,PrW	FWL	D	P/G	7/82
Campbell Soup Co.	Napoleon, Ohio	150[5]	250	Sat	C,PrW[6]	JBC	Com	P	8/82
Campbell Soup Co.	Salisbury, Md.	50	150	Sat	C,PrW	JBC	Com	P	11/82
Boise Cascade Corp.	Kenora, Ont., Can.	45	250	Sat	W,S	EPI	Com	P	10/77
A Ahlstrom Oy[2]	Pori, Finland	44	1200	970	Pt,W	AHL	Com	P	1/79
Oy Kyro Ab[2]	Kyruskoski, Finland	44	870	914	W,Pt	TAM	Com	P	5/81
House of Raeford	Rose Hill, N. C.	43	150	Sat	W,PI	YSI	Com	P	5/82
City of Kemijarvi	Kemijarvi, Finland	41[3]	232[3]	356[3]	Pt,W	TAM	Com	P	11/83
Central Soya Company	Marion, Ohio	40	200	Sat	C,Ng	JBC	Com	P	4/80
Undisclosed	Undisclosed	40[7]	120	Sat	C,Ng	JBC	Com	P	3/83
Tampella Ltd.[2]	Anjalankoski, Finland	40	1420	Sat	W,S,C	TAM	Com	P	11/82

TABLE E-3. EXISTING AND PLANNED MULTI-FUEL AND ALTERNATUVE FUEL AFBC UNITS[35] (Continued)

Plant Owner	Location	Steam Capacity, 1000 lb/hr	Steam Pressure psig	Steam Temperature °F	Design Fuel(s)	Manufacturer	Type of Project	Type of Financing	Commercial Service Date
City of Bolinas	Bolinas, Sweden	34[1]	175	375	R,W	GEN	Com	P	9/83
City of Landskrona	Landskrona, Sweden	34[1]	175	375	RDF	GEN	Com	P	8/83
City of Vastervik	Vastervik, Sweden	34[1]	175	375	R,W	GEN	Com	P	6/84
Woolcombers Ltd.	Bradford, Eng.	25	200	Sat	C,PrW	FWL	Com	P/G	8/82
Tobacco Processing	Brazil	25[5]	150	Sat	C,AL	JBC	Com	P	2/81
Undisclosed	Providence, R.I.	20[1]	300	Sat	Ng,C	JBC	Com	P	5/83
Lumber Mill	Crestview, Fla.	20	300	Sat	W,Ng	JBC	Com	P	3/81
IBM	Charlotte, N. C.	20	225	Sat	Ng,O,C	JBC	Com	P	7/80
Undisclosed	Erving, Mass.	20	150	Sat	C,O,Ng	JBC	Com	P	4/83
G.A. Serlachium Lielahti[2]	Tampere, Finland	19	653	842	S,W,Pt	TAM	Com	P	2/80
Hayward Tyler Pump Co.	Keighley, Eng.	10	125	Sat	C,Ng	JBC	Com	P	1/80
U.S. Department of HUD	Norfolk, Va.	10[3]	20[3]	200[3]	Wo,Ti	DFC[8]	Com	G	10/82
Tenneco Organics Ltd.	Avonmouth, Eng.	6	250	Sat	Wt,Wo	DFC	Com	P	6/80
Undisclosed	Rome, Italy	6	150	Sat	Wt,PrW	DFC	Com	P	NAv
Struthers Thermo-Flood	Winfield, Kan.	5	2650	660	C,L,Ck	SWC	T	P	10/81

FOOTNOTES, ABBREVIATIONS, AND MANUFACTURERS FOR TABLE E-3

Footnotes:

1. Two units installed.
2. Application for fluidized-bed boiler is steam production in a papermill.
3. Rating is in million Btu/hr; unit is a fluidized-bed hot-water boiler. Operating pressure and temperature are for hot water.
4. Rating is in million Btu/hr; hot combustion gas exiting fluidized-bed combustor flows to a conventional fired boiler.
5. Three units installed.
6. Also oil and natural gas.
7. Nine units installed.
8. In conjunction with International Boiler Works Co.

Abbreviations:

Ac--Anthracite culm
Al--Alcohol
Bc--Brown coal
C--Coal
Ch--Cotton hulls
Ck--Petroleum coke
CO--Carbon monoxide
Com--Commercial contract
Cww--Coal-washing wastes
D--Demonstration project
D/C--Demonstration/commercial project
G--Government financing
L--Lignite
NAv--Not available
Ng--Natural gas
O--Oil
Op--Olive pits
P--Private financing
P/G--Private/government financing
PL--Poultry litter
PrW--Process wastes
Pt--Peat
R--Refuse
RDF--Refuse-derived fuel
S--Sludge
Sat--Saturated
T--Test facility
Ti--Tires
W--Wood, woodwaste, wood byproducts
Wo--Waste oil
Wt--Waste tars

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FOOTNOTES, ABBREVIATIONS, AND MANUFACTURERS FOR TABLE E-3 (Continued)

Manufacturers:

AHL--Ahlstrom Oy
DBW--Deutsche Babcock Werke AG
DED--Dedert Corp., Thermal Processes Division
DFC--Deborah Fluidized Combustion Ltd.
DNL--Danka of Netherton Ltd.
EPI--Energy Products of Idaho
FEC--Fluidyne Engineering Corporation
FWC--Foster Wheeler Boiler Corporation
FWL--Foster Wheeler Power Products Ltd.
GEN--Generator Industrie AB
HIT--Babcock Hitachi KK
IHI--Ishikawajima-Harima Heavy Industries Co.
JBC--Johnston Boiler Co.
KEE--E. Keeler Co.
NEI--NEI Cochran Ltd.
SWC--Struthers Wells Corp.
TAM--Tampella Ltd.
YSI--York Shipley, Inc.